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Final Report
Contract CPA 22-69-78

FEASIBILITY STUDY OF NEW SULFUR OXIDE CONTROL PROCESSES FOR APPLICATION TO SMELTERS AND POWER PLANTS

Part II: The Wellman-Lord SO_2 Recovery Process for Application to Smelter Gases

Prepared for:

U.S. DEPARTMENT OF HEALTH, EDUCATION, AND WELFARE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION
DURHAM, NORTH CAROLINA

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Prepared for:

U.S. DEPARTMENT OF HEALTH, EDUCATION, AND WELFARE
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION
DURHAM, NORTH CAROLINA

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A block flow diagram was prepared.
Mass and volume flow balances were estimated. Engineering estimates of the capital investment and the total annual cost (including both fixed and variable charges) were prepared. These engineering estimates were applied to previously created model gas streams and the total annual cost for each was determined. From these annual costs, incremental production costs were estimated. The technical constraints on the application and operation were qualitatively appraised; the economic constraints were appraised (quantitatively, to the extent permitted by the data). The current state of development was assessed, identifying any technological deficiencies whose elimination might enhance applicability.

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FOREWORD

The final report for this study is presented in four separate and independent parts:

Part I: The Monsanto Cat-Ox Process for Application to Smelter Gases

Part II: The Wellman-Lord SO_2 Recovery Process for Application to Smelter Gases

Part III: The Monsanto Cat-Ox Process for Application to Power Plant Flue Gases

Part IV: The Wellman-Lord SO_2 Recovery Process for Application to Power Plant Flue Gases

Information for use in this study was supplied to Stanford Research Institute by Monsanto Company and Wellman-Lord, Inc. under terms of confidentiality agreements between the U.S. Department of Health, Education and Welfare, Stanford Research Institute, and each of the cooperating companies. In accordance with the agreements, Monsanto Company and Wellman-Lord, Inc. have reviewed and released the parts of the report dealing with their respective processes. The rights of prior review and release are designed solely to permit the cooperating companies to assure themselves that no proprietary or confidential data are being revealed; they are not intended to restrict Stanford Research Institute's rights and responsibilities to report its conclusions so long as there is no incidental disclosure of confidential information. Accordingly, the release of the reports by Monsanto and Wellman-Lord does not imply that these companies necessarily concur in all or any of the opinions, judgments, or interpretations of fact expressed by the author, who assumes sole responsibility for the report content.

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CPA 22-69-78.

I INTRODUCTION

Under the Systems Study for Control of Emissions -- Primary Nonferrous Smelting Industry (Contract No. PH 86-68-85), Arthur G. McKee & Company and its subcontractor, Stanford Research Institute, carried out evaluations of a number of sulfur oxide control processes as they might be applied to offgases from nonferrous smelting. To permit evaluation of the technical and economic feasibility of these control processes, a number of models of smelters were created. Stanford Research Institute carried out the studies necessary to determine the availability of markets for sulfur by-products open to smelters in various areas, and the allowable production costs that the smelters would have to attain in order to break even on the sulfur recovery operations.

The Division of Process Control Engineering of the National Air Pollution Control Administration (DPCE-NAPCA) desires to extend the usefulness of the foregoing study by adding to it technical and economic evaluations of new and potentially promising sulfur oxide control processes. It also wishes to evaluate the same new processes for application to power plants. Completion of these preliminary evaluations of the processes will help determine their potential commercial acceptability.

DPCE-NAPCA has a specific interest in at least two control processes being offered commercially, the Monsanto Cat-Ox process and the Wellman-Lord SO_2 Recovery process. However, both processes are proprietary, and, as a matter of policy, DPCE-NAPCA does not wish to obtain proprietary and confidential information on the processes. It does, nevertheless, wish to obtain evaluations in nonconfidential terms. Broadly, DPCE-NAPCA wishes to obtain estimates of the capital and annual costs of the control systems for each of the assumed applications, together with appraisals of the technical constraints on each process and of the current states of development of the processes.

Stanford Research Institute was requested by DPCE-NAPCA to carry out evaluations of the processes under the terms of confidentiality agreements between the Department of Health, Education, and Welfare, the owners of the proprietary processes, and Stanford Research Institute. SRI, acting as a disinterested third party, was to make analyses of the processes using information obtained from Monsanto Company and Wellman-Lord, Inc., and to report the results to DPCE-NAPCA without compromising any of the Monsanto or Wellman-Lord confidential data.

Because of the requirements of confidentiality, it is not permissible to describe certain features of the Cat-Ox and Wellman-Lord processes. The corresponding portions of the systems have had to be represented only in terms of their general functions, and SRI's evaluations of these portions has had to be presented in the form of conclusions without supporting data or reasoning. In other instances, the parts of the systems could be described in general, but specific details and design parameters could not be revealed.

Within the scope of the present project it would obviously have been impossible to inspect and evaluate independently all the company records and design data even had the cooperating companies been requested to permit this and had they acceded to the request. The author of this report, who also conducted the study, evaluated the information provided at his request, using his own knowledge and relevant data from the literature and other available sources. Whenever apparent discrepancies or uncertainties were noted in the information, efforts were made to secure verification or clarification from the companies. In instances where resolution of questions was not possible, or the information required proved to be simply unavailable, the author employed his best judgment.

Throughout the following sections of this report, information for which other sources are not specifically cited was generally obtained from the cooperating companies and accepted by the author either because it could be verified from other sources or because it appeared reasonable. In other instances, information or estimates were provided by

the companies that could not be verified independently or judged for reasonableness; in such cases, the companies have been specifically cited as the sources. In still other instances, the author did not accept the information or estimates provided and in some cases substituted his own; such cases have also been specifically noted.

The cooperating companies, at their own option and through substantial efforts, provided the basic capital cost estimates for the model control systems and the information for estimation of operating and maintenance costs. The author in this case acted as a reviewer rather than as an estimator. The estimates were checked for reasonableness and for possible errors or omissions. For some components and cost factors, the author modified the estimates, or substituted others of his own where he judged them to be more appropriate than those supplied to him. The author also prepared cost estimates for some auxiliary systems, using separate data sources.

By specification of the power plant and smelter models, and by review of the results, an effort was made to ensure that the cost estimates for both control systems were made on strictly comparable bases. Although it is unlikely that this objective has been met fully, the deviations are probably within the precision of the estimates themselves.

For convenience, and at the request of DPCE-NAPCA, this final report is presented in four separate and independent parts. This part deals only with the Wellman-Lord SO_2 Recovery process as applied to control of sulfur oxides in the offgases from copper, lead, and zinc smelters.

II OBJECTIVES

The objectives of the part of the study covered by this report are as follows:

1. To prepare a block flow diagram of the Wellman-Lord system showing its configuration and its relation to the smelting processes in which the sulfur oxides are generated.

2. To present estimated mass and volume flow balances for the Wellman-Lord system.

3. To prepare preliminary engineering estimates of the capital investment and the total annual cost (including both fixed and variable charges) for the Wellman-Lord system.

4. To apply the estimates prepared in (3) to the gas streams of the model smelters created in the previous Systems Study for Control of Emissions -- Primary Nonferrous Smelting Industry, and to determine the total annual cost for each model control system. From the estimates of total annual costs, secondary estimates are to be made of the corresponding incremental costs of producing the nonferrous metals, both on the gross basis (without allowance for by-product recovery credits) and on the net basis (with allowance for by-product recovery credits).

5. To make a qualitative appraisal of technical constraints on the application and operation of the control system.

6. To appraise (quantitatively, to the extent permitted by available data) the economic constraints on the application of the control system.

7. To assess the current state of development of the Wellman-Lord system, identifying any technological deficiencies whose elimination might enhance the applicability of the system to smelter gases.

The accomplishment of the objectives is subject to any restrictions that may be imposed under the terms of the confidentiality agreement between the Government, Wellman-Lord, Inc., and Stanford Research Institute.

III SUMMARY

The Wellman-Lord SO_2 Recovery process is a cyclic absorption-desorption process for producing concentrated sulfur dioxide from waste gas streams, such as power plant flue gas and smelter offgases. It is adaptable to use on gas streams containing sulfur dioxide at both low and relatively high concentrations.

The basic Wellman-Lord system is made up of two basically different parts: (1) the absorption section, in which the sulfur dioxide is removed from the flue gas, and (2) the chemical recovery section, in which the concentrated sulfur dioxide is recovered from the spent absorbent and the absorbent is regenerated for return to the absorption section. The absorbent is a solution of sodium sulfite. In the absorption cycle, the sulfite ion reacts with sulfur dioxide, forming bisulfite. In the recovery cycle, the reaction is reversed by application of heat, releasing sulfur dioxide and regenerating the sulfite.

The absorption step is conventional and has been employed commercially as well as studied experimentally. The regeneration step apparently represents an innovation; it is intended to reduce the amount of steam required where regeneration is accomplished by direct stripping of the bisulfite solution with steam. The Wellman-Lord regeneration (or chemical recovery) system appears to be capable of operating with a steam demand in the range of about one-half to one-third that required in a conventional steam stripping system. The regeneration equipment is relatively expensive, but not sufficiently so for the associated fixed charges to offset the advantage of the reduction in steam consumption. Since the regeneration system has not yet received patent protection, it cannot be described at this time.

The Wellman-Lord system treated in this study is essentially a conceptual design. Components of the system have been tested individually, but a complete, integrated system has not yet been operated. The regeneration system is basically different from that used in another version

of the process using potassium sulfite solution as the absorbent. The latter system, which was tested in a demonstration plant on power plant flue gases, was generally unsatisfactory. However, the Wellman-Lord system as designed for use with the sodium-base absorbent appears to be technically feasible.

It is essential that the gas entering the Wellman-Lord system be as free as possible from dust, fume, and vapor or gaseous contaminants, such as arsenic trioxide, hydrogen chloride, hydrogen fluoride, and sulfur trioxide. Solid particles in the absorbent solution may produce a variety of mechanical and chemical problems, while acidic gases and vapors will consume absorbent. In addition, a number of the contaminants may tend to catalyze the oxidation of the sulfite to sulfate. It was therefore specified by SRI as a model condition that the gas should be assumed to be cleaned to the same degree as it would be for use in a contact sulfuric acid plant.

In the Wellman-Lord process, as in all similar processes, some portion of the sulfite (or sulfur dioxide) is oxidized to sulfate, from which the sulfur dioxide cannot be regenerated. The sulfate must be purged from the system. The fraction of the sulfur dioxide that is oxidized has not been established. Wellman-Lord originally estimated that the amount would be very low, and provided no system for recovering the sodium base from the purge stream of absorbent solution, which was expected to be small. The author anticipated that the amount of sulfur dioxide oxidized would be much larger than that originally projected by Wellman-Lord, and assumed the amount to be 1.0 percent of that absorbed, although the actual percentage is unpredictable. He therefore assumed that the purged absorbent solution would be recausticized with lime to recover the sodium base.

The purging of the sulfate results in a loss of sulfur dioxide in the sodium sulfite present in the purge stream. If as much as 4 to 6 percent of the sulfur dioxide should be oxidized, a substantial fraction of the total sulfur dioxide will be lost in the purge, and a system for separating the sodium sulfate and sulfite will be needed.

The product of the Wellman-Lord system, concentrated sulfur dioxide, has only very limited markets. Hence, it will in general be necessary either to convert the sulfur dioxide to sulfuric acid in an auxiliary contact acid plant, or to reduce it to elemental sulfur in an auxiliary reduction plant. The markets for sulfuric acid are limited in the areas of the United States where there are the greatest number of smelters with uncontrolled emissions, and where elemental sulfur is a more desirable by-product.

In the model studies (using the smelter models created by Arthur G. McKee & Company), the Wellman-Lord process was applied conceptually in 12 control schemes producing sulfuric acid or elemental sulfur. Where sulfuric acid was to be the product, the Wellman-Lord system was applied only to gas streams containing 2 percent or less of sulfur dioxide, and conventional contact acid plants were applied to richer gases. Where elemental sulfur was to be the product, the Wellman-Lord process was applied to all the gas streams to concentrate the sulfur dioxide for subsequent reduction.

The limiting economic factor in the application of the Wellman-Lord process proved to be the capital and operating costs of the chemical recovery section, which dominated the total cost. The indicated costs of recovery of sulfur dioxide from lean gases were high, as with all other recovery processes. With richer gases, the recovery cost remained high because the size of the chemical recovery section was proportional to the quantity of sulfur dioxide regardless of the concentration in the gas treated. For all of the control schemes producing elemental sulfur, the cost of production exceeded by factors of two or more any realizable price that might be obtained for the sulfur. Of those control schemes producing sulfuric acid, only one attained a production cost that might be reasonably competitive in some smelter areas; in this scheme, sulfur dioxide was concentrated from reverberatory furnace gases in a copper smelter and fed to a conventional sulfuric acid plant operating on the roaster and converter gases.

For the various smelter models and control schemes, the gross incremental costs of producing the metals (before allowance for by-product credits) ranged as follows:

Copper: 1.5 to 3.6 cents/lb

Zinc: 1.1 to 2.3 cents/lb

Lead: 0.6 to 1.1 cents/lb

The by-product credits that might be realized would not in most cases reduce these costs by a large factor.

IV PROCEDURES

A. Models

The models of the hypothetical smelters, which are presented in Appendix A, are the same as those formulated and used in the previous study of smelter emission control by Arthur G. McKee & Company.[16] Some supplemental conditions, necessary to the specific analyses of the Cat-Ox and Wellman-Lord processes, were specified by SRI and are also given in Appendix A.

B. Cost Factors

The cost factors used in making the cost estimates for the control systems were also taken from the McKee report,[16] and are presented in Appendix B. Some supplemental factors needed specifically for the present study were specified by SRI.

The estimates of the prices that might be obtained for sulfur by-products were also taken from the McKee report.[16]

C. Preparation of Technical Data and Cost Estimates

Wellman-Lord, Inc. prepared capital cost estimates for gas absorbers of three different capacities and for chemical recovery sections of five different capacities, and also estimated the utilities and labor requirements. The author reviewed these estimates and accepted most of them. The cost breakdowns for individual components of the system were supplied by Wellman-Lord for review. Most of the components consisted of conventional items of chemical process equipment, and the costs were estimated by Wellman-Lord on the basis of quotations from vendors. The author surveyed these estimates for general reasonableness and consistency with the model specifications, but did not attempt further verification.

Some of the original estimates of system requirements or costs made

by Wellman-Lord were not accepted, and the author supplied his own estimates. Such instances are noted in the following sections of this report. The changes were discussed with Wellman-Lord, and differences were generally resolved.

The cost estimates for the gas cleaning system and the auxiliary sulfuric acid and sulfur dioxide reduction plants were prepared by the author (see Appendixes C, D, and E), from published data.

V PROCESS DESCRIPTION

The Wellman-Lord SO_2 Recovery process is a cyclic absorption-desorption process for producing concentrated sulfur dioxide from various concentrations in waste gas streams, such as power plant flue gas and smelter offgases. It is adaptable to use on gas streams containing sulfur dioxide at both low and relatively high concentrations. Its product, concentrated sulfur dioxide, has very limited markets, but can be used in sulfuric acid manufacture or be reduced to elemental sulfur. Therefore, the Wellman-Lord process potentially has two principal uses: (1) concentration of sulfur dioxide from dilute gas streams (those containing less than about 2 to 3 percent of sulfur dioxide) for use in sulfuric acid plants, and (2) concentration of sulfur dioxide from both dilute and rich gas streams for subsequent reduction to elemental sulfur. If sulfuric acid is the desired product, and the gas stream contains at least 3.5 to 4 percent of sulfur dioxide, it is practical to treat the gas directly in a contact sulfuric acid plant.[6]

The absorption step in the Wellman-Lord process cycle is conventional, and in closely related forms has been used commercially and been studied widely on an experimental basis by numerous investigators. Wellman-Lord, Inc. believes that some elements of the absorber design are novel and is seeking patent protection for them. Although the writer questions -- on purely technical grounds -- whether the design features are actually novel, it is not critical to the present evaluation of the basic process whether the claimed innovations in equipment are real or not.

The absorbent consists of a solution of sodium or potassium sulfite. In the absorption cycle, the sulfite ion reacts with sulfur dioxide as indicated forming bisulfite:

$$SO_3^{=} + SO_2 + H_2O = 2\ HSO_3^{-}$$

In the desorption cycle, the reaction is reversed by application of heat

releasing sulfur dioxide and regenerating the sulfite:

$$2\ HSO_3^- = SO_3^= + H_2O + SO_2$$

The absorption step has been employed commercially in the pulp and paper industry (usually with sodium and ammonia as bases) for production of cooking liquor;[1,5,19] in this case no regeneration step is employed. The complete cycle, including the desorption step, has been studied experimentally[13,14] and employed occasionally on a limited commercial scale.[13] However, the major economic limitation has been the quantity of steam required for the absorbent regeneration. In the conventional approach to regeneration,[14] the rich absorbent (containing a large fraction of bisulfite) is stripped of sulfur dioxide with steam in a countercurrent tower. At the bottom of the tower, where the lean stripped absorbent solution emerges, the bisulfite has been largely converted to sulfite and the back pressure of sulfur dioxide above the solution is low. The practical limits of regeneration of the absorbent are set by the required tower height and the quantity of steam.

Various attempts have been made to avoid the steam stripping of the absorbent. Johnstone,[15] for example, proposed introducing a chemical regeneration cycle. The Wellman-Lord process employs crystallization of salts from the absorbent as part of a procedure for reducing the steam requirements for regeneration. Various proposed forms of the process are described in Wellman-Lord patents.[3,17,18] One, which uses the potassium-base salts, is described broadly in a Belgian patent,[3] and more specifically in a paper by Watt.[23] It was studied on a pilot plant scale[22] and was later investigated in a demonstration plant.[23] The anticipated steam economy was not attained, and development has been shifted to another form of the process, which employs the sodium-base salts.

The sodium-base process, which is used as the basis for the present study, has a regeneration process that is essentially different from that employed where the potassium-base salts are used. It is the regeneration process that constitutes the apparently novel feature of the Wellman-Lord system. Applications have been made for patents, but patent protection

has not yet been received. Consequently, it is not permissible at this time to describe the basic regeneration process or the system for carrying it out. The corresponding portions of the regeneration system are therefore omitted from the flowsheet, Fig. 1.

The system considered in the present study has not been operated in a complete, integrated system, but experiments have been carried out with individual components. It is essentially a conceptual design backed by relevant experience with the potassium-base pilot plant and demonstration plant and by data from the tests of system components. A commercial plant using the process is to go into service in 1970 on the tail gas from a contact sulfuric acid plant.

The regeneration system devised by Wellman-Lord avoids the direct steam stripping of the absorbent in a stripping column, and reduces the steam requirement substantially below that of the stripper. During regeneration, the amount of sulfur dioxide recovered per unit quantity of absorbent circulated depends upon the concentration of the solution and upon the fraction of the salt that is initially bisulfite. Hence, attaining maximum steam economy (the least consumption of steam per unit quantity of sulfur dioxide recovered) requires that the rich absorbent leaving the absorption process be as concentrated as possible, and that the conversion of the sulfite to bisulfite also be as nearly complete as possible.

The upper limit of concentration of the sodium sulfite in the lean absorbent solution fed to the absorber is set by the solubility of the salt. As the sulfite is converted to the more soluble bisulfite by reaction with sulfur dioxide (the actual crystalline compound is the purosulfite instead of the bisulfite), the solution becomes less nearly saturated. Hence, a portion of the water in the solution can be evaporated into the gas stream passing through the tower. However, a practical limit on the amount of water removed in this manner is imposed by the need to avoid bringing the rich absorbent solution to saturation with respect to sodium pyrosulfite. If saturation is reached, plugging of parts of the absorber and absorbent outlet piping by pyrosulfite crystals

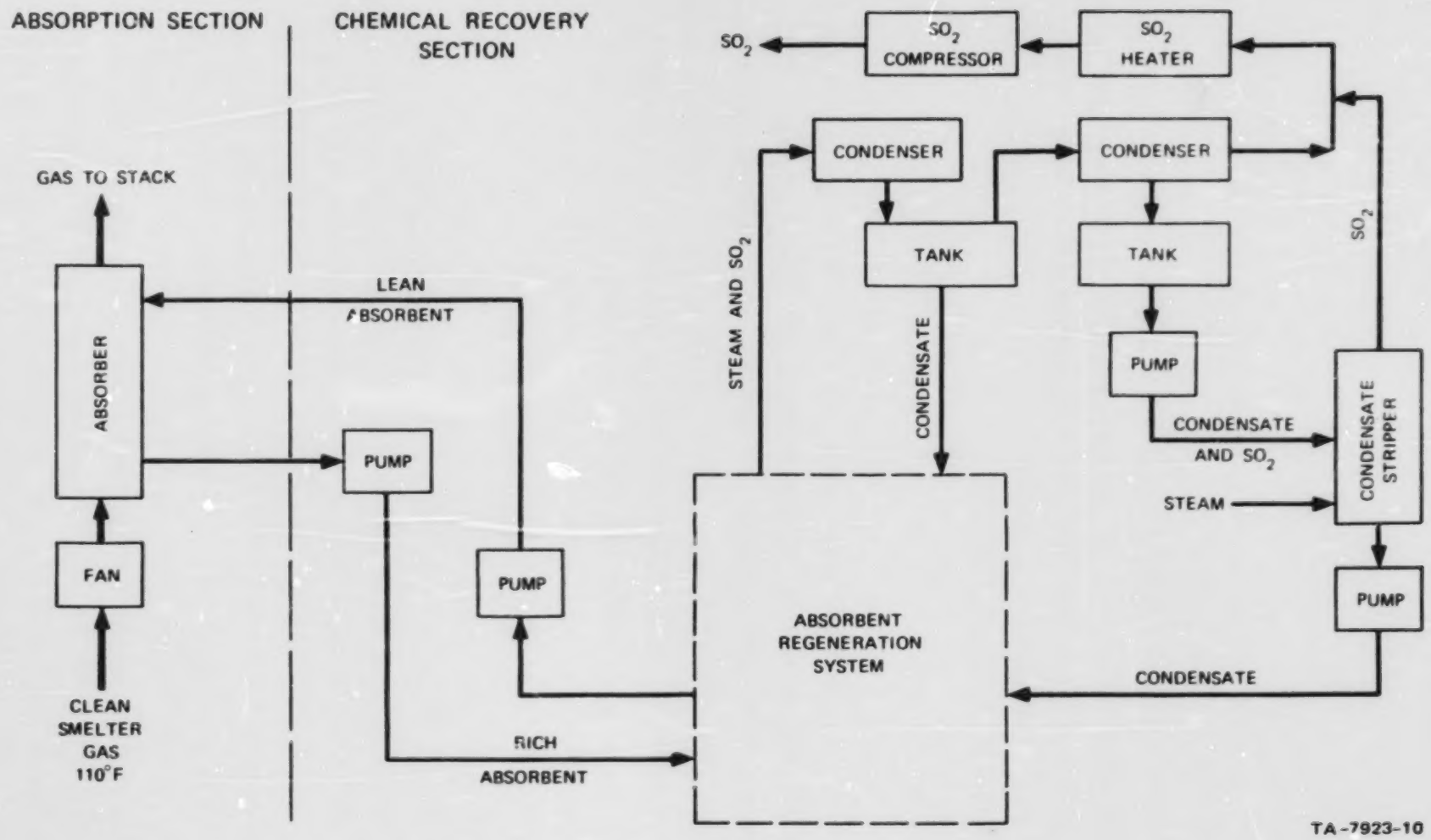

FIGURE 1 WELLMAN-LORD SO_2 RECOVERY SYSTEM FOR SMELTER GASES

may occur.

The circulation rate of the absorbent is set by the quantity of sulfur dioxide to be removed from the gas stream. If the gas being treated is relatively rich in sulfur dioxide (as, for instance, a smelter gas), it is possible in a countercurrent absorption tower to convert somewhat more of the sulfite to bisulfite than if the gas is more dilute; a higher back pressure of sulfur dioxide above the solution leaving the tower is permissible because the partial pressure of sulfur dioxide in the entering gas stream is also higher. However, the additional fractional conversion of the sulfite is not large,[14] and the gain in capacity of the absorbent is ignored in conservative design of a system. The sulfur dioxide capacity is therefore taken as constant regardless of the composition of the gas stream treated.

The steam economy of the system can be improved by staging the regeneration process, but the economic limits of this approach are set by the concomitant increase in capital investment. The optimum balance between operating costs and depreciation must therefore be determined.

In the Wellman-Lord process, as in all similar processes employing the sulfite-bisulfite absorption and recovery cycle (or even absorption without regeneration), some portion of the sulfite is oxidized to sulfate, from which the sulfur dioxide cannot be regenerated. The sulfate must therefore be purged from the system. The fraction of the sulfite that may be oxidized in the Wellman-Lord system has not yet been established; it may, in fact, vary widely depending upon such factors as the oxygen content of the gas stream, the temperature, and the presence of contaminants that may act as oxidation catalysts. Wellman-Lord adds hydroquinone to the absorbent solution as an oxidation inhibitor, but the effectiveness of this measure is questionable. This point is discussed below.

The basic Wellman-Lord system is made up of two basically different parts, the absorption section and the chemical recovery section (see Fig. 1). The absorption section is composed only of the absorber and the fan. Its size, and hence its cost, is dependent almost solely upon

the volume of gas handled, with only a minor cost component related to the concentration of the sulfur dioxide in the gas. The size and cost of the chemical recovery section are dependent upon the total quantity of sulfur dioxide recovered. Since the quantity of sulfur dioxide absorbed per unit quantity of absorbent is constant, the total volume of absorbent circulated is directly proportional to the total quantity of sulfur dioxide.

Removal of particulate matter from the gas before the latter enters the absorber is a critical aspect of the operation of the Wellman-Lord process. (It will evidently be equally critical to any other cyclic process for concentration of sulfur dioxide.) Particulate matter entering the absorption system may produce mechanical problems in the chemical recovery section (e.g., plugging lines and wearing pumps) and may interfere in other ways with the operation of the system. It must be purged from the system by filtration or centrifuging and may carry with it occluded absorbent. If the filtered particulate is washed to recover the absorbent, the resulting dilution of the solution increases the steam requirements for the regeneration process. The particulate matter may also include elements that will act as catalysts for the oxidation of the sulfite.

The gas stream may contain, in addition to sulfur dioxide, other gases and vapors that are deleterious to the system. Examples are sulfur trioxide, hydrogen chloride, hydrogen fluoride, and arsenic trioxide. A wet scrubbing system is therefore an essential part of the preliminary gas cleaning system, both for removing soluble gaseous contaminants and for cooling and humidifying the gas stream. Sulfur trioxide (and sulfuric acid mist), hydrogen chloride, and hydrogen fluoride all form sodium salts, consuming the absorbent base and requiring purging from the system. It was specified by SRI (Apperdix A) that the gas entering the system should be assumed to: (1) have been purified in a gas cleaning system typical of those used with contact sulfuric acid plants, (2) contain residual contaminants in concentrations at least as low as are specified by Donovan and Stuber,[7] and (3) be at 110°F and saturated with

water vapor. The capital and operating costs of the gas cleaning system are charged against sulfur oxide recovery, but were estimated separately by SRI from literature data (see Appendix C) and are presented as separate items.

The clean smelter gas is forced up through the countercurrent absorbing tower where it contacts the absorbent solution on a series of trays and the sulfur dioxide is absorbed. The treated gas leaving the top tray passes through an entrainment separator where entrained droplets of the absorbent solution are removed, and then is discharged to the stack. Arrangements for gas-liquid contacting in the tower vary somewhat depending upon the concentration of the sulfur dioxide in the gas. For sulfur dioxide concentrations under 2.5 percent the liquid throughput is relatively low, and special provisions are made to ensure good gas-liquid contact. These provisions (not indicated in Figure 1) make the capital cost somewhat higher than that for an absorber for a richer gas, in which the liquid throughput is higher and no extra measures are necessary to ensure effective gas-liquid contacting.

A portion of the sulfuric acid mist remaining in the cleaned smelter gas will be collected in the absorber. The amount is not readily predictable, but the author assumed 90 percent.

The gas pressure drop across the absorber is approximately 12 inches of water. The fan is sized and powered to supply the draft for movement of the gas through the gas cleaning system, which has an additional pressure drop of 14 inches of water. The fan and large circulating pumps of the Wellman-Lord system are driven by backpressure steam turbines. The exhaust steam (at 50 psig) is sufficient to supply the requirements of the chemical recovery section. Only small pumps and other equipment are electrically driven.

The rich absorbent solution leaving the bottom of the absorbing tower is pumped to the chemical recovery section, where the sulfur dioxide is released from the bisulfite solution and the sulfite is reformed. A mixture of the sulfur dioxide and steam leaves the system and goes first to the main condenser, which is operated at about 200°F. Most of the

steam is condensed, but the temperature is high enough that little of the sulfur dioxide is absorbed in the condensate, which can be returned to the system as makeup water. The hot, wet sulfur dioxide stream leaving the receiving tank of the main condenser enters the second condenser, where it is cooled to about 120°F. Most of the remaining water vapor condenses, but the condensate contains a high concentration of dissolved sulfur dioxide. This second condensate stream must therefore be fed to the condensate stripper, where the sulfur dioxide is recovered by stripping with a small amount of steam. The overhead from the stripper joins the main gas stream from the second condenser receiving tank. At this point the main stream of sulfur dioxide gas is at about 120°F and saturated with water vapor, which then constitutes less than 12 percent by volume of the gas stream. The stripper overhead contains a much higher concentration of water vapor, but is only a small fraction of the main gas stream. The wet product sulfur dioxide is heated to prevent subsequent condensation of water vapor, then compressed to about 8 psig for delivery to an adjacent plant for subsequent manufacture of sulfuric acid or reduction to elemental sulfur.

To prevent the buildup of an excessive sodium sulfate concentration in the absorbent, a purge stream of the lean absorbent solution is withdrawn. In the original design supplied by Wellman-Lord for this study, no provision was made for recovery of the sodium base from the absorbent purge stream because it was believed that the amount of sulfite (or sulfur dioxide) oxidized to sulfate would be very low. The author believes that the amount of sulfur dioxide oxidized will probably be much higher and that recovery of the purged base will be an economic necessity. If the amount of sulfur dioxide oxidized does not exceed perhaps 1 to 2 percent of that absorbed, it may be feasible to recausticize the sodium sulfate-sodium sulfite mixture with lime, precipitating the sulfate and sulfite. The recovered sodium hydroxide solution may have to be carbonated to precipitate calcium ions remaining in solution. If calcium ions remain in the solution, they may eventually produce scaling problems in the recovery system. As a guesstimate, the capitalized cost of operating such a recovery system might be about $12 to $15 per ton of sodium hydroxide recovered,

including the cost of lime, makeup sodium hydroxide, operation, and depreciation of equipment. The higher figure, $15/ton, is assumed in this study.

In the purging of sulfate, a portion of the sulfur dioxide is lost because of purging of the associated sulfite. The author calculated the average sulfur dioxide loss to be about 6.5 percent of that absorbed from the flue gas, assuming that 1.0 percent is oxidized and that 90 percent of the sulfuric acid mist in the clean smelter gas is collected in the absorber to form additional sulfate. If the amount of sulfur dioxide oxidized should be as much as perhaps 4 to 6 percent of that absorbed, the sodium hydroxide could still be recovered by the recausticization process, but the amount of sulfur dioxide lost in purging would be a substantial fraction of the total. In such case, there is a need for a system for separating and purging the sodium sulfate that will not result in loss of the sulfite.

There will be a gradual buildup of other contaminants (such as chlorides, fluorides, and solid particles) in the absorbent, originating from the residuals in the clean smelter gas. These contaminants will also have to be purged from the system along with the sulfate.

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VI PROCESS DATA AND COST ESTIMATES

A. Sulfur Oxides Recovery

In the smelter models (Appendix A), the distribution of the emitted sulfur oxides between sulfur dioxide and sulfur trioxide is in most cases not given. Where the concentration of sulfur trioxide in the gas is not specified, the concentration of sulfur dioxide is calculated on the assumption that all of the sulfur is in the form of sulfur dioxide (see Table A-1). In fact, a variable quantity of sulfur trioxide (equivalent to a few percent of the total sulfur emission) will be present in all cases, and will be removed in the gas cleaning system that precedes the sulfur dioxide recovery system. However, for the sake of simplification, it was assumed in this study that all of the sulfur was in the form of sulfur dioxide and was treated in the sulfur dioxide control system. The inputs and outputs of sulfur from the recovery systems were calculated on this basis.

In the McKee study,[16] different collection efficiencies for sulfur dioxide were assumed for the various control systems modeled, as appeared appropriate for the systems considered. In the present study, a uniform standard of performance was set for both of the recovery systems (Cat-Ox and Wellman-Lord) that were treated (see Appendix A). Both systems appeared to be capable of attaining or exceeding the efficiencies specified, at least within the ranges of input sulfur dioxide concentrations for which their application was appropriate. The calculated collections of sulfur dioxide were based on attainment of the specified efficiencies.

The overall efficiency of sulfur dioxide recovery with the Wellman-Lord system is reduced by the inefficiencies of the auxiliary sulfuric acid or sulfur dioxide reduction plant (see Appendixes D and E). The losses from the auxiliary plant could be reduced by applying a Wellman-Lord absorber to the tail gas and returning the spent absorbent to the same chemical

recovery section that serves the main gas absorption system. However, no such additional absorbers were included in the control system models.

B. Gas Cleaning

The costs (capital and operating) of preliminary gas cleaning (see Appendix C) were taken to be the same for both the Cat-Ox and Wellman-Lord systems. However, the costs for the gas cleaning and sulfur dioxide recovery systems are presented separately in this report, and are combined only in final totals.

The capital cost for the gas cleaning system does not include that for the fan, which is assigned to the sulfur dioxide recovery system. However, the operating cost for the gas cleaning system does include the cost of electrical power consumed by the fan that moves the gas through the gas cleaning system.

The maximum allowable concentrations of impurities in the cleaned gas (Appendix A and Reference 7) are apparently higher, in at least a few instances, than the levels that are favored by some other authorities or that are attained in some installations. For sulfuric acid mist, Donovan and Stuber[7] give an upper limit of 0.022 grain/std cu ft, which is assumed in this study, whereas Carter[4] favors a maximum of 0.0012 grain/cu ft. Heinrich and Anderson[12] report attainment of residual acid mist and arsenic concentrations as low as 0.0000022 grain/cu ft with a wet electrostatic precipitator.

C. Cost Estimates for Wellman-Lord System

To permit relatively rapid estimation of costs for the various model control schemes, curves of capital and annual costs were prepared for the Wellman-Lord system (Figs. 2 through 5). Separate curves were derived for the absorption section (based on gas flow) and for the chemical recovery section (based on sulfur dioxide input). Similar curves were prepared for the gas cleaning system (Appendix C), the auxiliary sulfuric acid plant (Appendix D), and the sulfur dioxide reduction plant (Appendix E). Costs

for the various system components were then read off from the appropriate curves to derive cost estimates for the control schemes.

The capital cost for the absorption section (Fig. 2) was taken from Wellman-Lord estimates, except that the author added an allowance for ductwork connecting the outlet of the absorber to the stack that was assumed to be an existing part of the smelter. The capital cost for the chemical recovery section (Fig. 4) was taken from the Wellman-Lord estimates without change. Utilities and labor requirements for the absorption section (Table I) and chemical recovery section (Table IV) were based on Wellman-Lord estimates. The ratios of sodium hydroxide and hydroquinone makeup to sodium sulfate purged were also supplied by Wellman-Lord, but the author made his own estimates of the amount of sulfate formed (see Section V). The original estimate of sodium hydroxide entrainment loss from the absorber was tenfold higher than that adopted by the author, who believes that the values given in Table I should be reasonable unless reentrainment should take place in the eliminator.

Losses of sodium hydroxide due to entrainment and reaction with sulfuric acid mist are proportional to gas flow and were charged against the absorption section. Those due to oxidation of the sulfur dioxide are proportional to the quantity of sulfur dioxide and were charged to the chemical recovery section, even though the oxidation actually takes place in the absorber (except for possible autoxidation of the sulfite).

All of the steam used to drive the turbine-powered fans is later used for solvent regeneration in the chemical recovery section, and the entire steam cost was charged to the latter (Table IV). Since the steam condensate is returned to the boiler plant, the charge for the steam is reduced by a credit for the value of the condensate.

The absorber sections for use on the more dilute gases require electrical drives for equipment not employed where richer gases are treated (Table I). The chemical recovery sections (particularly the smaller ones) for use on plants treating dilute gases have a relatively high proportion of electrical drives (Table IV). Where richer gases are

being treated and steam requirements for regeneration are higher, it is practical to drive large rotating equipment in the chemical recovery section with turbines, since the exhaust steam can be used for regeneration.

The computations of annual costs for the absorption and chemical recovery sections are presented in Tables II, III, and V. The effect of the sulfur dioxide concentration in the gas on the annual cost of the chemical recovery section was so small that only a single cost curve was drawn in Fig. 5.

D. Cost Estimates for Model Control Schemes

The gas flow rates and sulfur dioxide emissions from the model smelters are summarized in Table VI. Because the copper converter gas flow rate varies widely, the rate used in designing gas handling equipment was assumed to be higher than the average (see Appendix A). The performances of the Wellman-Lord systems on the various smelter gas streams are presented in Table VII.

The smelter model control schemes treated are described in Table VIII, and the cost estimates are given in Tables IX through XX. All the capital and annual costs were taken from the curves in this report except those for the conventional contact sulfuric acid plants used in Control Schemes CA-1 and CB-1, which were taken from curves in Reference 16. Conversion in these acid plants was assumed to be 97%.

In Figs. 6 through 13 the changes in the unit production costs of the metals, resulting from the application of the control schemes to the model smelters, are shown graphically as functions of the net selling price of the sulfur by-product at the smelter. (The quantities of metal produced by the model smelters are given in Appendix A.) The net selling price is equal to the gross selling price less sales cost and overhead, and will, of course, be dependent upon the particular location of the plant.

Table I

UTILITIES, LABOR, AND CHEMICAL REQUIREMENTS
OF WELLMAN-LORD ABSORBERS

Gas Flow Rate ACFM (110°F, 1 atm, sat'd)	Electric Power[1] (kw)	Sodium Hydroxide Makeup (Entrainment Loss) (lb/hr)	Sodium Hydroxide Makeup (Acid Mist Reaction) (lb/hr)	Hydroquinone (lb/hr)	Labor $\left(\frac{\text{man-hr}}{\text{day}}\right)$
37,000	6.6	0.37	24.3	0.05	12
229,000	19.8	2.29	150	0.27	
580,000	52.6	5.80	381	0.70	
Cost Factor	1.0¢/kwh	$65.00/ton[2]	$15.00/ton[3]	82¢/lb	$3.75/hr

[1] Only with SO_2 under 2.5%.

[2] Purchased sodium hydroxide.

[3] Recovered sodium hydroxide.

Table II

CAPITAL AND ANNUAL COSTS OF WELLMAN-LORD ABSORBERS
(SO_2 Less than 2.5%)

Item	Cost Basis	Gas Flow Rate-ACFM (110°F, 1 atm, sat'd)		
		37,000	229,000	580,000
Capital Cost	$	196,000	600,000	1,060,000
Annual Cost	$/yr			
A. Depreciation		13,070	40,000	70,670
B. Direct Operating Cost				
1. Labor		14,850	14,850	14,850
2. Supervision		—	--	--
3. Payroll benefits		3,710	3,710	3,710
4. Maintenance materials		5,880	18,000	31,800
5. Factory supplies		980	3,000	5,300
6. Electricity		520	1,570	4,170
7. Chemicals		1,860	11,250	28,670
C. Indirect Costs				
1. Controllable		10,370	16,430	23,330
2. Noncontrollable		5,880	18,000	31,800
Total Annual Cost	$/yr	57,120	126,810	214,300

Table III

CAPITAL AND ANNUAL COSTS OF WELLMAN-LORD ABSORBERS
(SO_2 greater than 2.5%)

Item	Cost Basis	Gas Flow Rate-ACFM (110°F, 1 atm, sat'd)		
		37,000	229,000	580,000
Capital Cost	$	186,000	580,000	1,030,000
Annual Cost	$/yr			
A. Depreciation		12,400	38,670	68,670
B. Direct Operating Cost				
1. Labor		14,850	14,850	14,850
2. Supervision		--	--	--
3. Payroll benefits		3,710	3,710	3,710
4. Maintenance materials		5,580	17,400	30,900
5. Factory supplies		930	2,900	5,150
6. Electricity		--	--	--
7. Chemicals		1,860	11,250	28,670
C. Indirect Costs				
1. Controllable		10,220	16,130	22,880
2. Noncontrollable		5,580	17,400	30,900
Total Annual Cost	$/yr	55,130	122,310	205,730

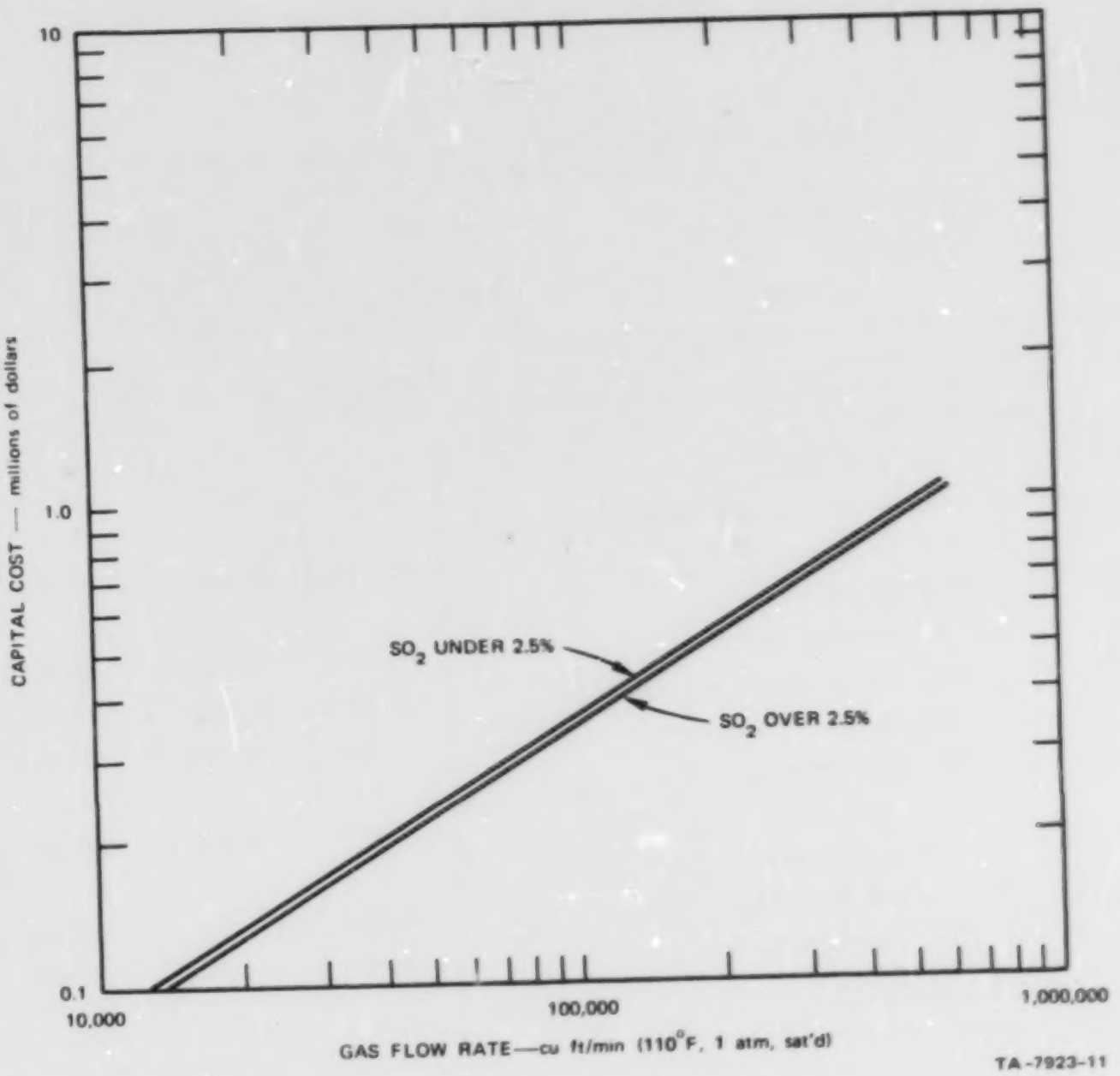

FIGURE 2 CAPITAL COSTS OF WELLMAN-LORD ABSORBERS

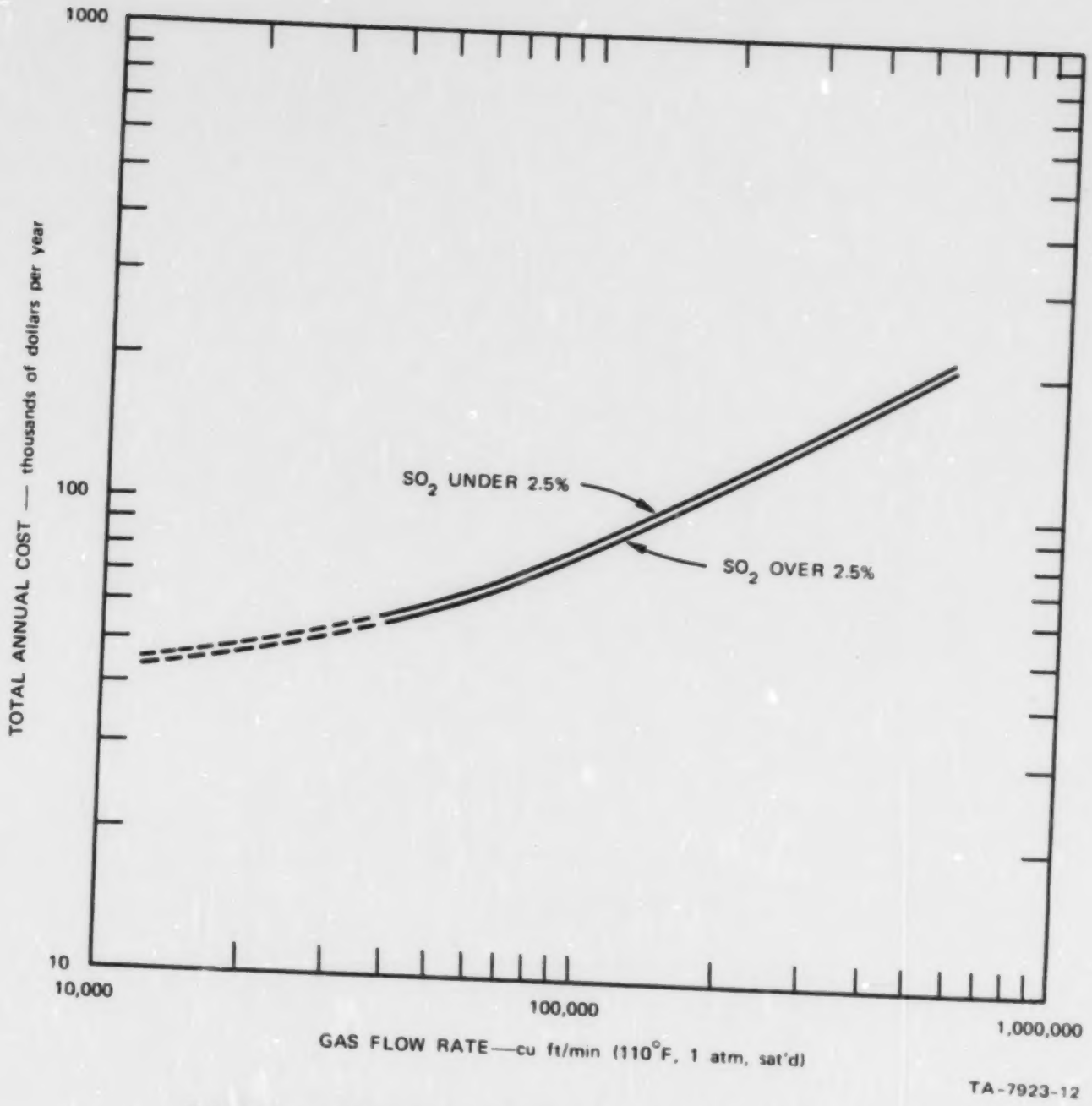

FIGURE 3 ANNUAL COSTS OF WELLMAN-LORD ABSORBERS

Table IV

UTILITIES, LABOR, AND CHEMICAL REQUIREMENTS OF
WELLMAN-LORD CHEMICAL RECOVERY SYSTEMS

Sulfur Dioxide Input (lb/hr)	Steam (lb/hr)	Electric Power (kw)		Process Water (lb/hr)[1]	Cooling Water Makeup (gpm)	Sodium Hydroxide (lb/hr)[2]	Hydroquinone (lb/hr)	Operating Labor $\left(\frac{\text{man-hr}}{\text{day}}\right)$[4]	Supervision $\left(\frac{\text{man-hr}}{\text{day}}\right)$
		SO_2> 1.5%	SO_2= 0.5%						
6,000	35,100	104	318	3,510	40	466	0.84	56	12
12,000	70,200	190	348	7,020	80	932	1.68		
24,000	140,400	309	349	14,000	160	1,870	3.37		
35,000	204,800	438	438	20,500	240	2,720	4.90		
49,000	286,700	557	557	28,700	330	3,810	6.86		
Cost Factor	75¢/1000 lb	1¢/kwh		5¢/1000 lb	2¢/1000 gal	$15/ton[3]	82¢/lb	$3.75/hr	$4.75/hr

[1] Steam condensate.

[2] Based on oxidation of 1.0 percent of input sulfur dioxide.

[3] Recovered sodium hydroxide.

[4] Two operators per shift plus one day laborer.

Table V

CAPITAL AND ANNUAL COSTS OF
WELLMAN-LORD CHEMICAL RECOVERY SYSTEMS

Item	Cost Basis	Sulfur Dioxide Input--lb/hr				
		6,000	12,000	24,000	35,000	49,000
Capital Cost	$	1,590,000	2,290,000	3,300,000	4,000,000	4,800,000
Annual Cost	$/yr					
A. Depreciation		106,000	152,670	220,000	266,670	320,000
B. Direct Operating Cost						
1. Labor		69,300	69,300	69,300	69,300	69,300
2. Supervision		18,810	18,810	18,810	18,810	18,810
3. Payroll benefits		22,030	22,030	22,030	22,030	22,030
4. Maintenance materials		47,700	68,700	99,000	120,000	144,000
5. Factory supplies		7,950	11,450	16,500	20,000	24,000
6. Steam		208,490	416,990	833,980	1,216,500	1,703,000
7. Electricity						
(a) SO_2 = 0.5%		25,190	27,560	27,640	34,690	44,110
(b) SO_2 > 1.5%		8,240	15,050	24,470	34,690	44,110
8. Process water		1,390	2,780	5,540	8,120	11,370
9. Cooling water makeup		380	760	1,520	2,280	3,140
10. Chemicals		33,130	66,270	132,960	193,390	270,860
C. Indirect Costs						
1. Controllable		67,900	78,400	93,360	104,060	116,060
2. Noncontrollable		47,700	68,700	99,000	120,000	144,000
Total Annual Cost	$/yr					
(a) SO_2 = 0.5%		655,970	1,004,420	1,639,840	2,195,850	2,890,680
(b) SO_2 > 1.5%		639,020	991,910	1,636,670	2,195,850	2,890,680

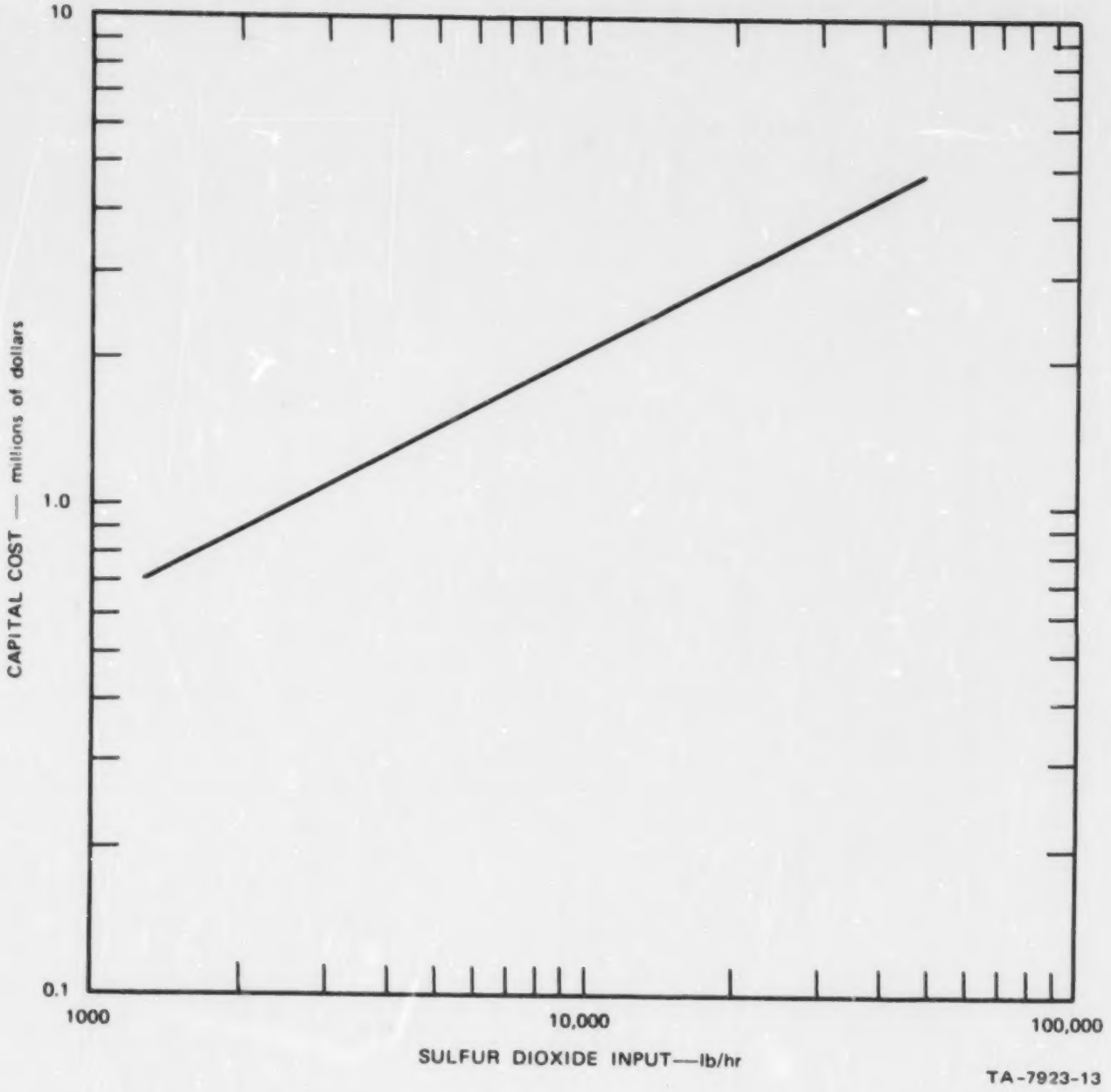

FIGURE 4 CAPITAL COSTS OF WELLMAN-LORD CHEMICAL RECOVERY SECTIONS

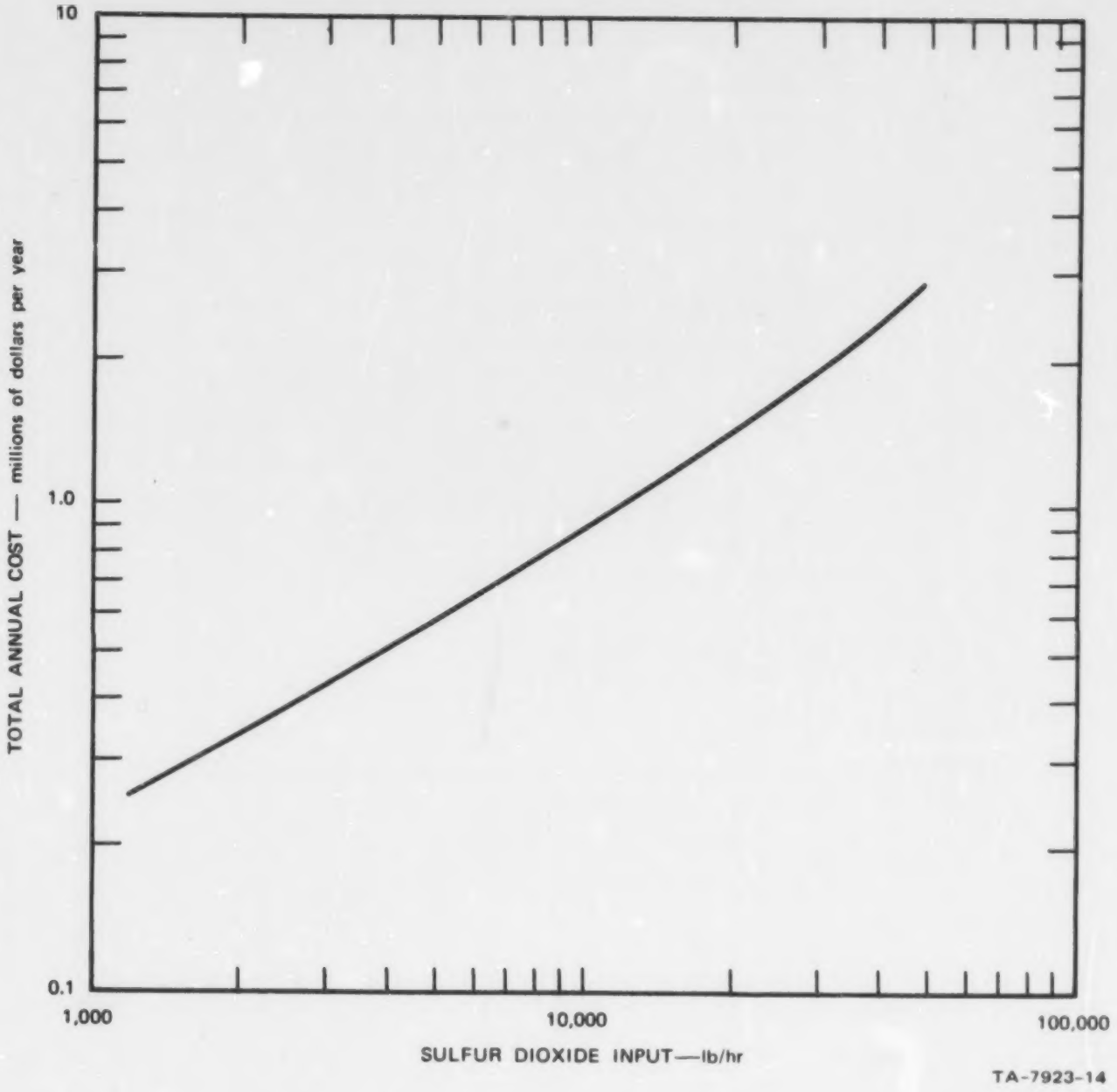

FIGURE 5 ANNUAL COSTS OF WELLMAN-LORD CHEMICAL RECOVERY SECTIONS

Table VI

GAS FLOW RATES AND SULFUR DIOXIDE EMISSIONS
FROM MODEL SMELTERS

Smelter	Model	Process Unit	Sulfur Dioxide Concentration (%)	Gas Flow Rate CFM (110°F, 1 atm, sat'd)			Sulfur Dioxide Emission (lb/hr)		
				Small	Medium	Large	Small	Medium	Large
Copper	A	Reverberatory Furnace	1.89	81,000	162,000	243,000	14,100	28,200	42,300
		Converters	3.82	75,100	150,200	225,300	26,480	52,970	79,450
		1		102,900	195,300	292,900			
	B	Roaster	11.2	29,800	59,600	89,400	30,630	61,270	91,900
		Reverberatory Furnace	0.91	48,800	97,700	146,600	4,100	8,200	12,300
		Converters	3.78	67,300	134,600	201,800	23,600	47,200	70,800
		1		87,500	165,600	248,200			
		Converters + Roaster	6.05	97,100	194,200	291,200	54,230	108,470	162,700
		2		117,200	225,200	337,600			
Zinc	A	Roaster	7.1	22,400	44,800	67,100	14,930	29,870	44,800
	B	Roaster	0.9	154,400			12,850		
		Sinter Plant	0.5	55,500			2,570		
	C	Roaster	0.8		409,600			30,300	
	D	Sinter-Roaster	2.0		166,900			30,870	
Lead	A	Sinter Plant	5.0	12,400	24,700	49,500	5,720	11,430	22,870

[1] Design basis for converters.

[2] Design basis for converters + roaster.

Table VII

RECOVERY OF SULFUR DIOXIDE FROM SMELTER
GASES BY WELLMAN-LORD SYSTEM

Smelter	Model	Process Unit	Sulfur Dioxide Absorption Efficiency (%)	Sulfur Dioxide Absorbed (lb/hr)			Sulfur Dioxide Recovered (lb/hr)[1]		
				Small	Medium	Large	Small	Medium	Large
Copper	A	Reverberatory Furnace	95	13,400	26,790	40,190	12,500	25,000	37,600
		Converters	98	25,950	51,910	77,860	24,300	48,500	72,800
	B	Roaster	98	30,020	60,040	90,060	28,100	56,100	84,200
		Reverberatory Furnace	94.51	3,870	7,750	11,600	3,620	7,250	10,800
		Converters	98	23,130	46,260	69,380	21,600	43,300	64,900
		Converters + Roaster	98	53,150	106,300	159,500	49,700	99,400	149,000
Zinc	A	Roaster	98	14,630	29,270	43,900	13,700	27,400	41,000
	B	Roaster	94.45	12,140			11,400		
		Sinter Plant	90	2,310			2,160		
	C	Roaster	93.75		28,400			26,600	
	D	Sinter-Roaster	95		29,300			27,400	
Lead	A	Sinter Plant	98	5,600	11,200	22,400	5,240	10,500	20,900

[1] Loss of sulfur dioxide assumed to be 6.5 percent of amount absorbed.

Table VIII

DESCRIPTION OF SMELTER MODEL CONTROL SCHEMES

Smelter Model	Control Scheme No.	Overall SO_2 Recovery (%)	Description
Copper A	CA-1	95.4	1. Contact sulfuric acid plant on converter gas. 2. Wellman-Lord recovery plant on reverberatory furnace gas. Concentrated SO_2 to contact acid plant on converter gas.
	CA-2	90.3	1. Individual Wellman-Lord absorbers on converter gas and reverberatory furnace gas. 2. Single Wellman-Lord chemical recovery section. Concentrated SO_2 to SO_2 reduction plant.
Copper B	CB-1	96.6	1. Contact sulfuric acid plant on mixed roaster and converter gas. 2. Wellman-Lord recovery plant on reverberatory furnace gas. Concentrated SO_2 to contact acid plant on roaster and converter gas.
	CB-2	91.0	1. Individual Wellman-Lord absorbers on reverberatory furnace gas and on mixed roaster and converter gas. 2. Single Wellman-Lord chemical recovery section. Concentrated SO_2 to SO_2 reduction plant.
Zinc A	ZA-1	91.3	Wellman-Lord recovery plant on roaster gas. Concentrated SO_2 to SO_2 reduction plant.

(continued)

Table VIII (concluded)

Smelter Model	Control Scheme No.	Overall SO_2 Recovery (%)	Description
Zinc B	ZB-1	92.0	1. Individual Wellman-Lord absorbers on roaster gas and sinter plant gas. 2. Single Wellman-Lord chemical recovery section. Concentrated SO_2 to contact sulfuric acid plant.
	ZB-2	87.3	1. Individual Wellman-Lord absorbers on roaster gas and sinter plant gas. 2. Single Wellman-Lord chemical recovery section. Concentrated SO_2 to SO_2 reduction plant.
Zinc C	ZC-1	92.0	Wellman-Lord recovery plant on roaster gas. Concentrated SO_2 to contact sulfuric acid plant.
	ZC-2	87.3	Wellman-Lord recovery plant on roaster gas. Concentrated SO_2 to SO_2 reduction plant.
Zinc D	ZD-1	93.2	Wellman-Lord recovery plant on sinter-roaster gas. Concentrated SO_2 to contact sulfuric acid plant.
	ZD-2	88.5	Wellman-Lord recovery plant on sinter-roaster gas. Concentrated SO_2 to SO_2 reduction plant.
Lead A	LA-1	91.3	Wellman-Lord recovery plant on sinter plant gas. Concentrated SO_2 to SO_2 reduction plant.

Table IX

COPPER SMELTER MODEL A--COSTS OF CONTROL SCHEME CA-1

Item	Cost or Credit		
	Small Plant	Medium Plant	Large Plant
Capital Investment			
1. Gas Cleaning System	$ 1,830,000	$ 2,950,000	$ 3,900,000
2. Wellman-Lord System			
(a) Absorption	319,000	490,000	630,000
(b) Chemical recovery	2,420,000	3,500,000	4,300,000
3. Contact Acid Plant	5,500,000	11,000,000	16,500,000
Total	$10,069,000	$17,940,000	$25,330,000
Annual Cost			
1. Gas Cleaning System	$ 358,000	$ 583,000	$ 780,000
2. Wellman-Lord System			
(a) Absorption	75,000	105,000	132,000
(b) Chemical recovery	1,090,000	1,760,000	2,420,000
3. Contact Acid Plant	1,400,000	2,800,000	4,200,000
Total	$ 2,923,000	$ 5,248,000	$ 7,532,000
Sulfuric Acid Produced (100% basis)	229,200 tons/yr	458,400 tons/yr	688,100 tons/yr
Gross Costs Before Credits			
1. Production Cost of Acid	$12.75/ton	$11.45/ton	$10.95/ton
2. Incremental Cost of Producing Metal	1.93¢/lb	1.73¢/lb	1.65¢/lb

Table X

COPPER SMELTER MODEL A-- COSTS OF CONTROL SCHEME CA-2

Item	Cost or Credit		
	Small Plant	Medium Plant	Large Plant
Capital Investment			
1. Gas Cleaning System			
(a) Reverberatory furnace	$ 1,830,000	$ 2,950,000	$ 3,900,000
(b) Converters	2,160,000	3,330,000	4,400,000
2. Wellman-Lord Absorbers			
(a) Reverberatory furnace	319,000	490,000	630,000
(b) Converters	355,000	525,000	675,000
3. Wellman-Lord Chemical Recovery	4,260,000	8,520,000	12,780,000
4. SO_2 Reduction Plant	1,560,000	2,340,000	2,950,000
Total	$10,484,000	$18,155,000	$25,335,000
Annual Cost			
1. Gas Cleaning System			
(a) Reverberatory furnace	$ 358,000	$ 583,000	$ 780,000
(b) Converters	423,000	665,000	900,000
2. Wellman-Lord Absorbers			
(a) Reverberatory furnace	75,000	105,000	132,000
(b) Converters	81,000	112,000	140,000
3. Wellman-Lord Chemical Recovery	2,380,000	4,760,000	7,140,000
4. SO_2 Reduction Plant	730,000	1,230,000	1,700,000
Total	$ 4,047,000	$ 7,455,000	$10,792,000
Elemental Sulfur Produced	67,500 short tons/yr	134,800 short tons/yr	202,500 short tons/yr
Gross Costs Before Credits			
1. Production Cost of Sulfur	$59.96/short ton	$55.30/short ton	$53.29/short ton
2. Incremental Cost of Producing Metal	2.67¢/lb	2.45¢/lb	2.37¢/lb

Table XI

COPPER SMELTER MODEL B--COSTS OF CONTROL SCHEME CB-1

Item	Cost or Credit		
	Small Plant	Medium Plant	Large Plant
Capital Investment			
1. Gas Cleaning System	$1,280,000	$ 2,080,000	$ 2,750,000
2. Wellman-Lord System			
(a) Absorption	233,000	358,000	460,000
(b) Chemical recovery	1,260,000	1,820,000	2,250,000
3. Contact Acid Plant	7,100,000	14,200,000	21,300,000
Total	$9,873,000	$18,458,000	$26,760,000
Annual Cost			
1. Gas Cleaning System	$ 256,000	$ 408,000	$ 540,000
2. Wellman-Lord System			
(a) Absorption	62,000	82,000	100,000
(b) Chemical recovery	500,000	765,000	985,000
3. Contact Acid Plant	1,700,000	3,400,000	5,100,000
Total	$2,518,000	$ 4,655,000	$ 6,725,000
Sulfuric Acid Produced (100% basis)	339,900 tons/yr	679,800 tons/yr	1,020,000 tons/yr
Gross Costs Before Credits			
1. Production Cost of Acid	$7.41/ton	$6.85/ton	$6.59/ton
2. Incremental Cost of Producing Metal	1.66¢/lb	1.53¢/lb	1.47¢/lb

Table XII

COPPER SMELTER MODEL B--COSTS OF CONTROL SCHEME CB-2

Item	Cost or Credit		
	Small Plant	Medium Plant	Large Plant
Capital Investment			
1. Gas Cleaning System			
(a) Reverberatory furnace	$ 1,280,000	$ 2,080,000	$ 2,750,000
(b) Roaster and converters	2,350,000	3,690,000	4,850,000
2. Wellman-Lord Absorbers			
(a) Reverberatory furnace	233,000	358,000	460,000
(b) Roaster and converters	380,000	575,000	740,000
3. Wellman-Lord Chemical Recovery	7,200,000	12,600,000	17,800,000
4. SO_2 Reduction Plant	1,930,000	2,900,000	3,650,000
Total	$13,373,000	$22,203,000	$30,250,000
Annual Cost			
1. Gas Cleaning System			
(a) Reverberatory furnace	$ 256,000	$ 408,000	$ 540,000
(b) Roaster and converters	460,000	740,000	1,000,000
2. Wellman-Lord Absorbers			
(a) Reverberatory furnace	62,000	82,000	100,000
(b) Roaster and converters	86,000	121,000	152,000
3. Wellman-Lord Chemical Recovery	3,680,000	6,900,000	10,200,000
4. SO_2 Reduction Plant	960,000	1,650,000	2,300,000
Total	$ 5,504,000	$ 9,901,000	$14,292,000
Elemental Sulfur Produced	97,800 short tons/yr	195,600 short tons/yr	293,100 short tons/yr
Gross Costs Before Credits			
1. Production Cost of Sulfur	$56.28/short ton	$50.62/short ton	$48.76/short ton
2. Incremental Cost of Producing Metal	3.63¢/lb	3.26¢/lb	3.13¢/lb

Table XIII

ZINC SMELTER MODEL A
COSTS OF CONTROL SCHEME ZA-1

Item	Cost or Credit		
	Small Plant	Medium Plant	Large Plant
Capital Investment			
1. Gas Cleaning System	$ 760,000	$1,220,000	$1,610,000
2. Wellman-Lord System			
(a) Absorption	137,000	210,000	270,000
(b) Chemical recovery	2,530,000	3,650,000	4,520,000
3. SO_2 Reduction Plant	870,000	1,310,000	1,650,000
Total	$4,297,000	$6,390,000	$8,050,000
Annual Cost			
1. Gas Cleaning System	$ 159,000	$ 242,000	$ 315,000
2. Wellman-Lord System			
(a) Absorption	49,000	58,000	67,000
(b) Chemical recovery	1,150,000	1,870,000	2,600,000
3. SO_2 Reduction Plant	400,000	600,000	790,000
Total	$1,758,000	$2,770,000	$3,772,000
Elemental Sulfur Produced	25,100 short tons/yr	50,300 short tons/yr	75,200 short tons/yr
Gross Costs Before Credits			
1. Production Cost of Sulfur	$70.04/short ton	$55.07/short ton	$50.16/short ton
2. Incremental Cost of Producing Metal	1.55¢/lb	1.23¢/lb	1.11¢/lb

Table XIV

ZINC SMELTER MODEL B
COSTS OF CONTROL SCHEME ZB-1

Item	Cost or Credit
Capital Investment	
1. Gas Cleaning Systems	
(a) Roaster	$2,850,000
(b) Sinter plant	1,410,000
2. Wellman-Lord Absorbers	
(a) Roaster	472,000
(b) Sinter plant	252,000
3. Wellman-Lord Chemical Recovery	2,520,000
4. Contact Acid Plant	645,000
Total	$8,149,000
Annual Cost	
1. Gas Cleaning Systems	
(a) Roaster	$ 560,000
(b) Sinter plant	278,000
2. Wellman-Lord Absorbers	
(a) Roaster	102,000
(b) Sinter plant	65,000
3. Wellman-Lord Chemical Recovery	1,140,000
4. Contact Acid Plant	212,000
Total	$2,357,000
Sulfuric Acid Produced (100% basis)	80,550 tons/yr
Gross Costs Before Credits	
1. Production Cost of Acid	$29.26/ton
2. Incremental Cost of Producing Metal	2.08¢/lb

Table XV

ZINC SMELTER MODEL B
COSTS OF CONTROL SCHEME ZB-2

Item	Cost or Credit
Capital Investment	
1. Gas Cleaning System	
(a) Roaster	$2,850,000
(b) Sinter plant	1,410,000
2. Wellman-Lord Absorbers	
(a) Roaster	472,000
(b) Sinter plant	252,000
3. Wellman-Lord Chemical Recovery	2,520,000
4. SO_2 Reduction Plant	860,000
Total	$8,364,000
Annual Cost	
1. Gas Cleaning System	
(a) Roaster	$ 560,000
(b) Sinter plant	278,000
2. Wellman-Lord Absorbers	
(a) Roaster	102,000
(b) Sinter plant	65,000
3. Wellman-Lord Chemical Recovery	1,140,000
4. SO_2 Reduction Plant	400,000
Total	$2,545,000
Elemental Sulfur Produced	24,900 short tons/yr
Gross Costs Before Credits	
1. Production Cost of Sulfur	$102/short ton
2. Incremental Cost of Producing Metal	2.25¢/lb

Table XVI

ZINC SMELTER MODEL C
COSTS OF CONTROL SCHEME ZC-1

Item	Cost or Credit
Capital Investment	
1. Gas Cleaning System	$ 5,550,000
2. Wellman-Lord System	
(a) Absorption	860,000
(b) Chemical recovery	3,600,000
3. Contact Acid Plant	990,000
Total	$11,000,000
Annual Cost	
1. Gas Cleaning System	$ 1,150,000
2. Wellman-Lord System	
(a) Absorption	176,000
(b) Chemical recovery	1,840,000
3. Contact Acid Plant	304,000
Total	$ 3,470,000
Sulfuric Acid Produced (100% basis)	158,000 tons/yr
Gross Costs Before Credits	
1. Production Cost of Acid	$21.96/ton
2. Incremental Cost of Producing Metal	1.54¢/lb

Table XVII

ZINC SMELTER MODEL C
COSTS OF CONTROL SCHEME ZC-2

Item	Cost or Credit
Capital Investment	
1. Gas Cleaning System	$ 5,550,000
2. Wellman-Lord System	
(a) Absorption	860,000
(b) Chemical recovery	3,600,000
3. SO_2 Reduction Plant	1,270,000
Total	$11,280,000
Annual Cost	
1. Gas Cleaning System	$ 1,150,000
2. Wellman-Lord System	
(a) Absorption	176,000
(b) Chemical recovery	1,840,000
3. SO_2 Reduction Plant	585,000
Total	$ 3,751,000
Elemental Sulfur Produced	48,800 short tons/yr
Gross Costs Before Credits	
1. Production Cost of Sulfur	$76.86/short ton
2. Incremental Cost of Producing Metal	1.66¢/lb

Table XVIII

ZINC SMELTER MODEL D
COSTS OF CONTROL SCHEME ZD-1

Item	Cost or Credit
Capital Investment	
1. Gas Cleaning System	$3,000,000
2. Wellman-Lord System	
(a) Absorption	500,000
(b) Chemical recovery	3,650,000
3. Contact Acid Plant	1,010,000
Total	$8,160,000
Annual Cost	
1. Gas Cleaning System	$ 595,000
2. Wellman-Lord System	
(a) Absorption	107,000
(b) Chemical recovery	1,870,000
3. Contact Acid Plant	310,000
Total	$2,882,000
Sulfuric Acid Produced (100% basis)	162,800 tons/yr
Gross Costs Before Credits	
1. Production Cost of Acid	$17.70/ton
2. Incremental Cost of Producing Metal	1.28¢/lb

Table XIX

ZINC SMELTER MODEL D
COSTS OF CONTROL SCHEME ZD-2

Item	Cost or Credit
Capital Investment	
1. Gas Cleaning System	$3,000,000
2. Wellman-Lord System	
(a) Absorption	500,000
(b) Chemical recovery	3,650,000
3. SO_2 Reduction Plant	1,310,000
Total	$8,460,000
Annual Cost	
1. Gas Cleaning System	$ 595,000
2. Wellman-Lord System	
(a) Absorption	107,000
(b) Chemical recovery	1,870,000
3. SO_2 Reduction Plant	600,000
Total	$3,172,000
Elemental Sulfur Produced	50,300 short tons/yr
Gross Costs Before Credits	
1. Production Cost of Sulfur	$63.06/short ton
2. Incremental Cost of Producing Metal	1.40¢/lb

Table XX

LEAD SMELTER MODEL A--COSTS OF CONTROL SCHEME LA-1

Item	Cost or Credit		
	Small Plant	Medium Plant	Large Plant
Capital Investment			
1. Gas Cleaning System	$ 500,000	$ 810,000	$1,310,000
2. Weilman-Lord System			
(a) Absorption	95,000	147,000	225,000
(b) Chemical recovery	1,530,000	2,200,000	3,180,000
3. SO_2 Reduction Plant	500,000	740,000	1,120,000
Total	$2,625,000	$3,897,000	$5,835,000
Annual Cost			
1. Gas Cleaning System	$ 115,000	$ 168,000	$ 259,000
2. Weilman-Lord System			
(a) Absorption	44,000	50,000	60,000
(b) Chemical recovery	625,000	960,000	1,530,000
3. SO_2 Reduction Plant	260,000	350,000	528,000
Total	$1,044,000	$1,528,000	$2,377,000
Elemental Sulfur Produced	9,610 short tons/yr	19,300 short tons/yr	38,300 short tons/yr
Gross Costs Before Credits			
1. Production Cost of Sulfur	$109/short ton	$79.1?/short ton	$62.06/short ton
2. Incremental Cost of Producing Metal	1.11¢/lb	0.81¢/lb	0.63¢/lb

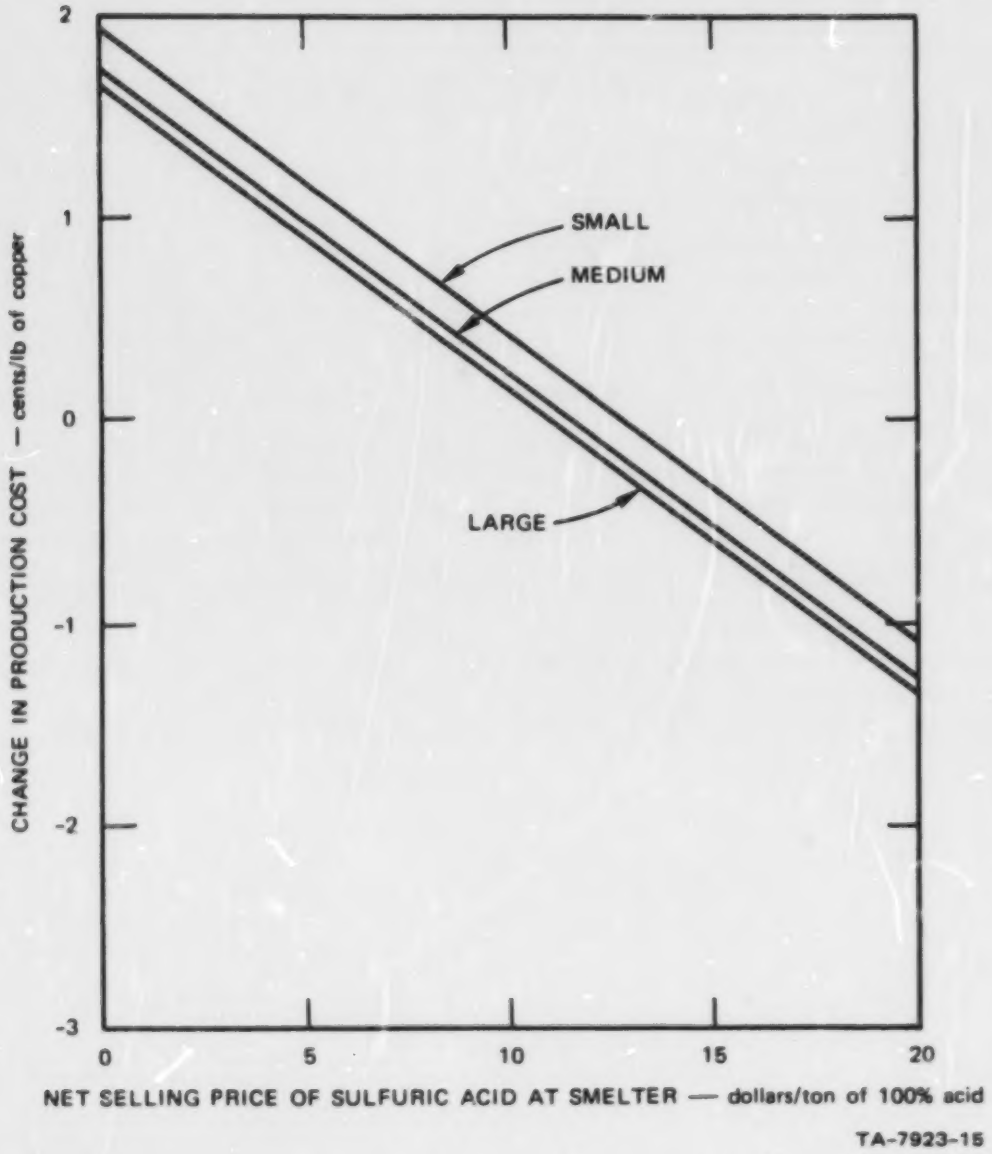

FIGURE 6 MODEL A COPPER SMELTERS: CONTROL SCHEME CA-1—EFFECT ON COPPER PRODUCTION COSTS

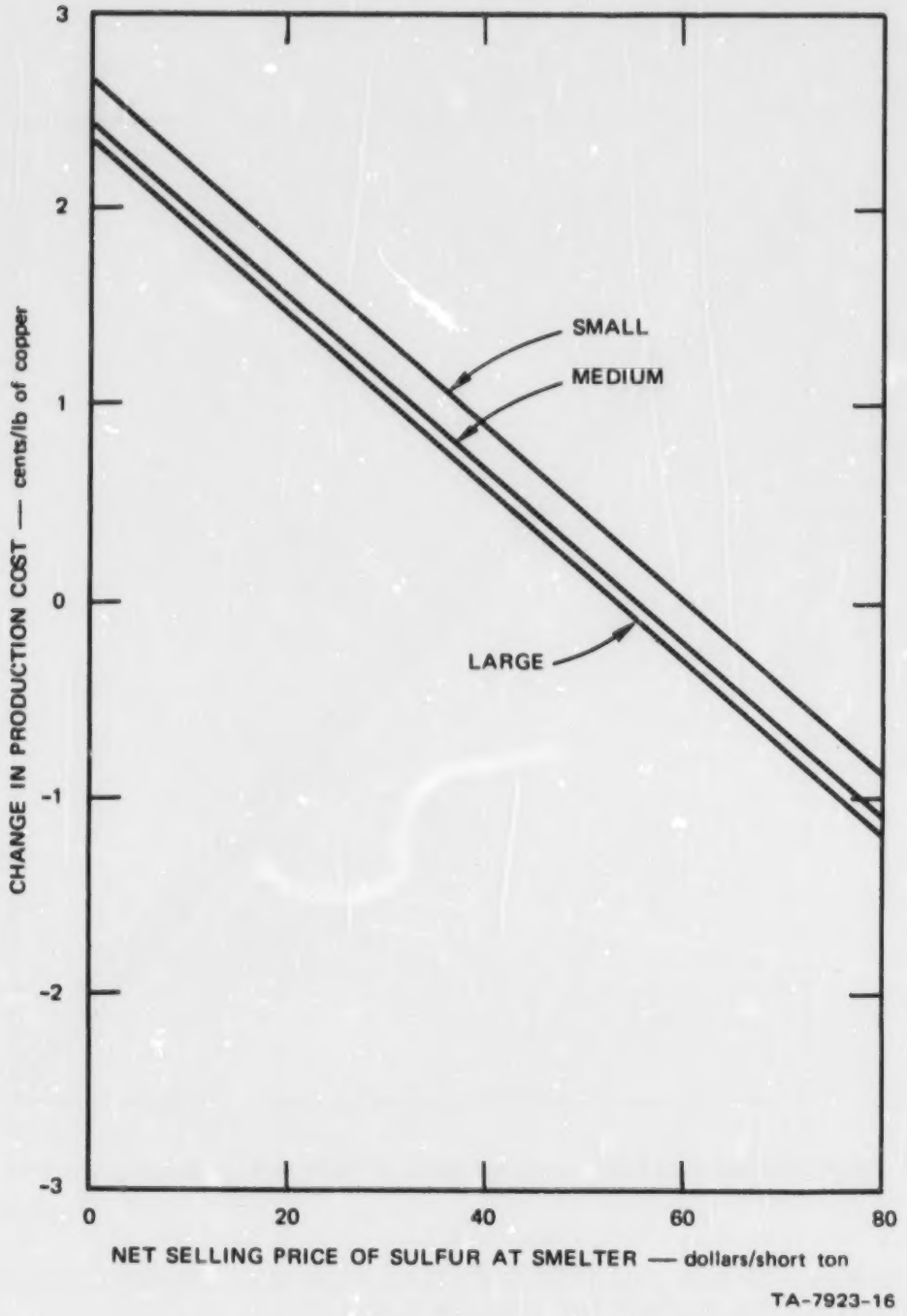

FIGURE 7 MODEL A COPPER SMELTERS: CONTROL SCHEME CA-2—EFFECT ON COPPER PRODUCTION COSTS

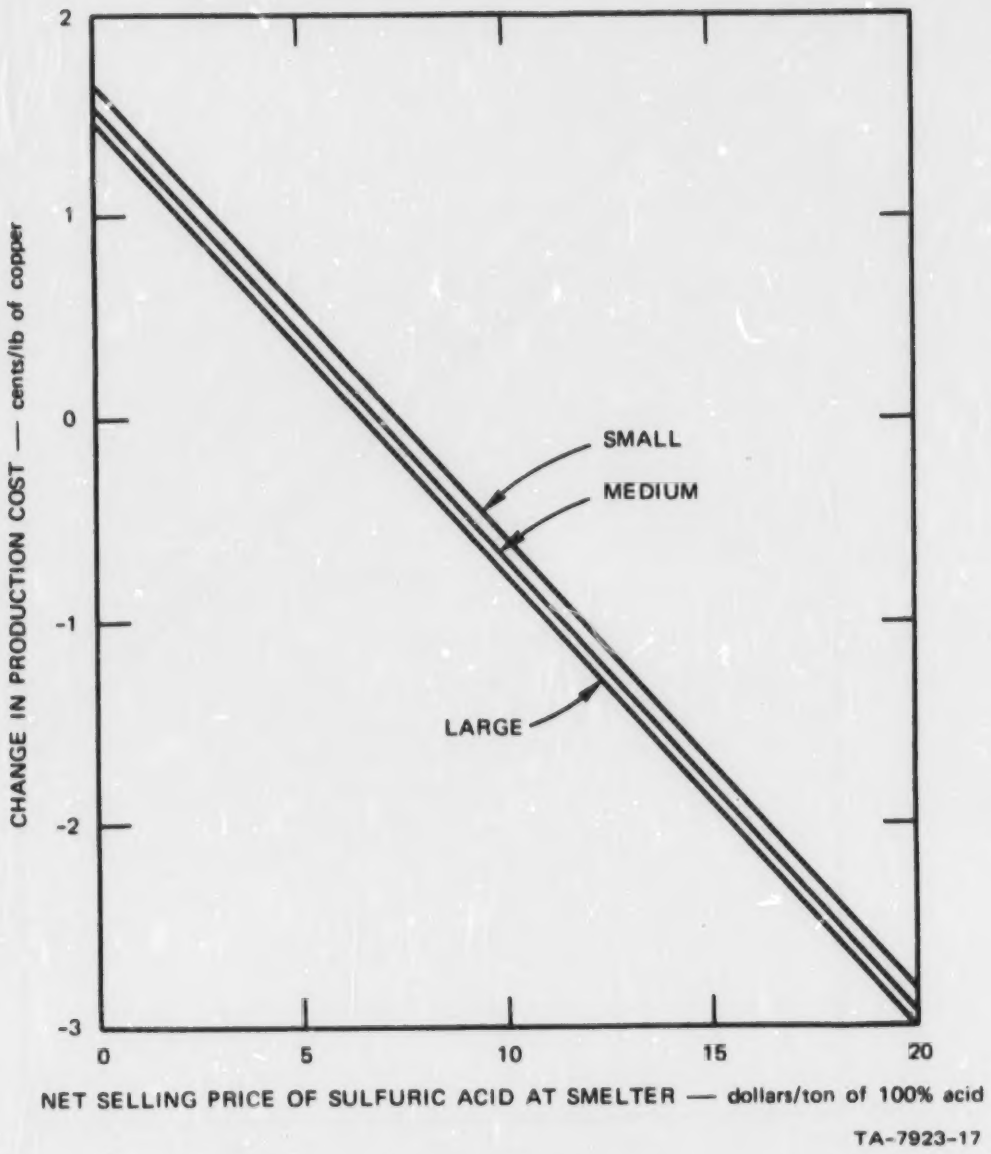

FIGURE 8 MODEL B COPPER SMELTERS: CONTROL SCHEME CB-1—EFFECT ON COPPER PRODUCTION COSTS

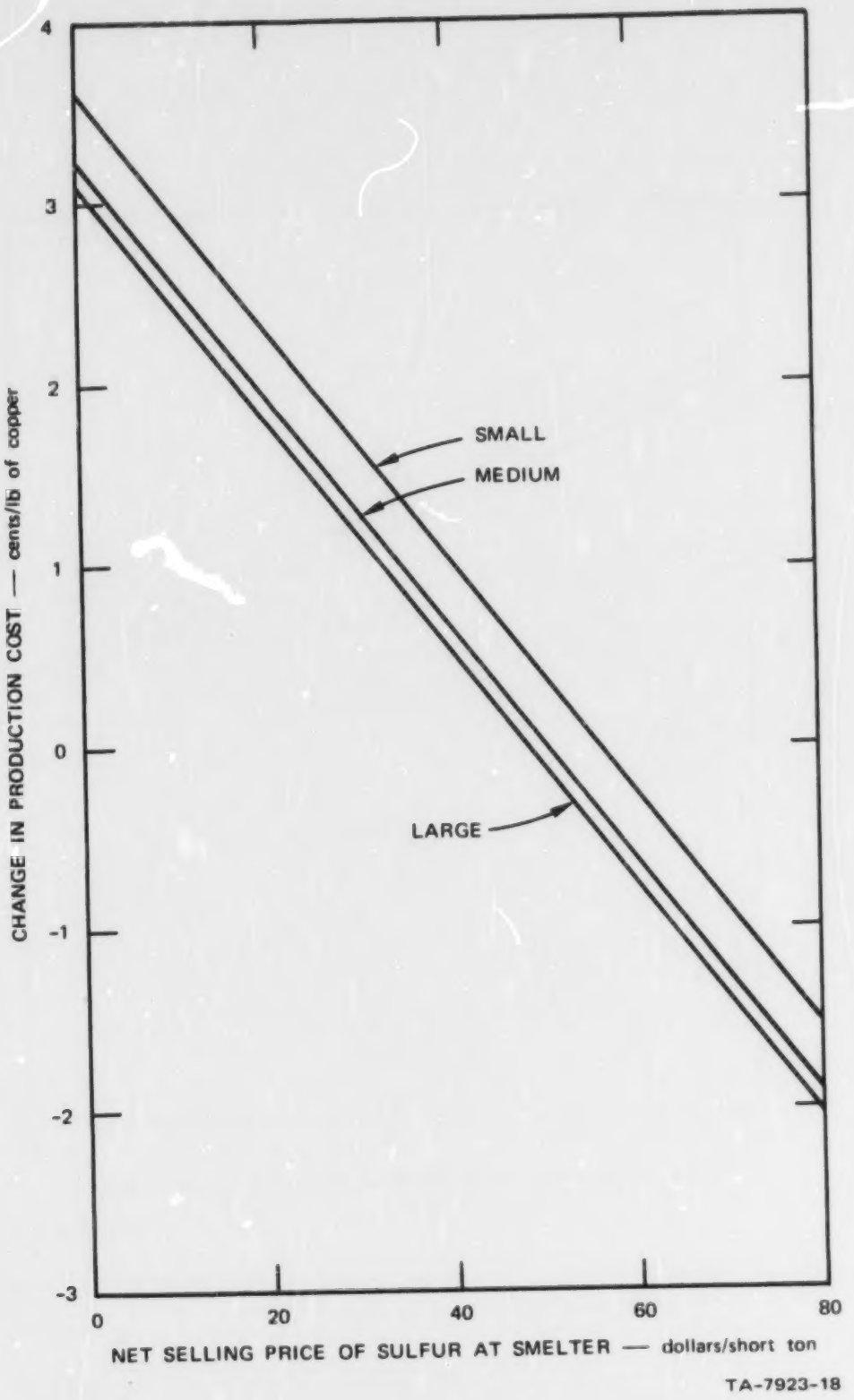

FIGURE 9 MODEL B COPPER SMELTERS: CONTROL SCHEME CB-2—EFFECT ON COPPER PRODUCTION COSTS

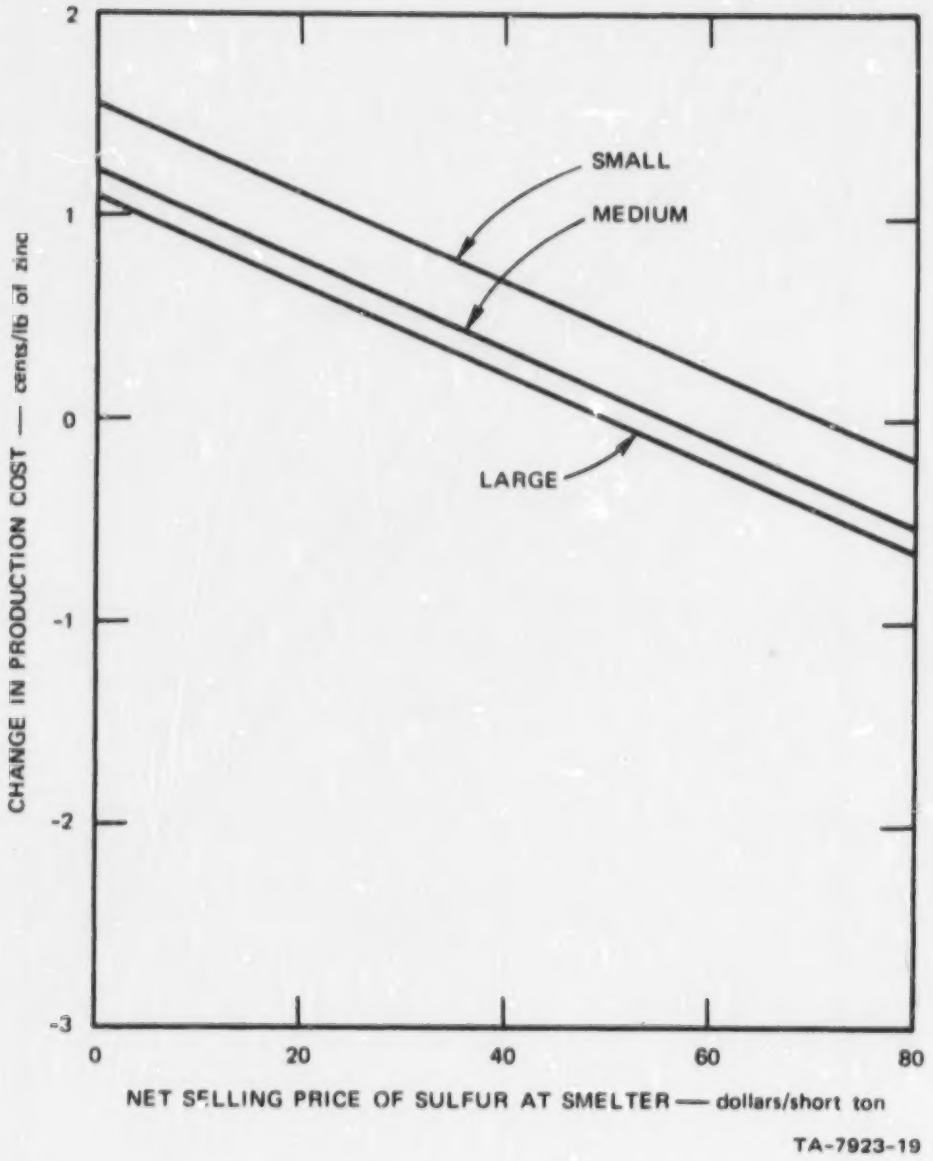

FIGURE 10 MODEL A ZINC SMELTERS: CONTROL SCHEME ZA-1—EFFECT ON ZINC PRODUCTION COSTS

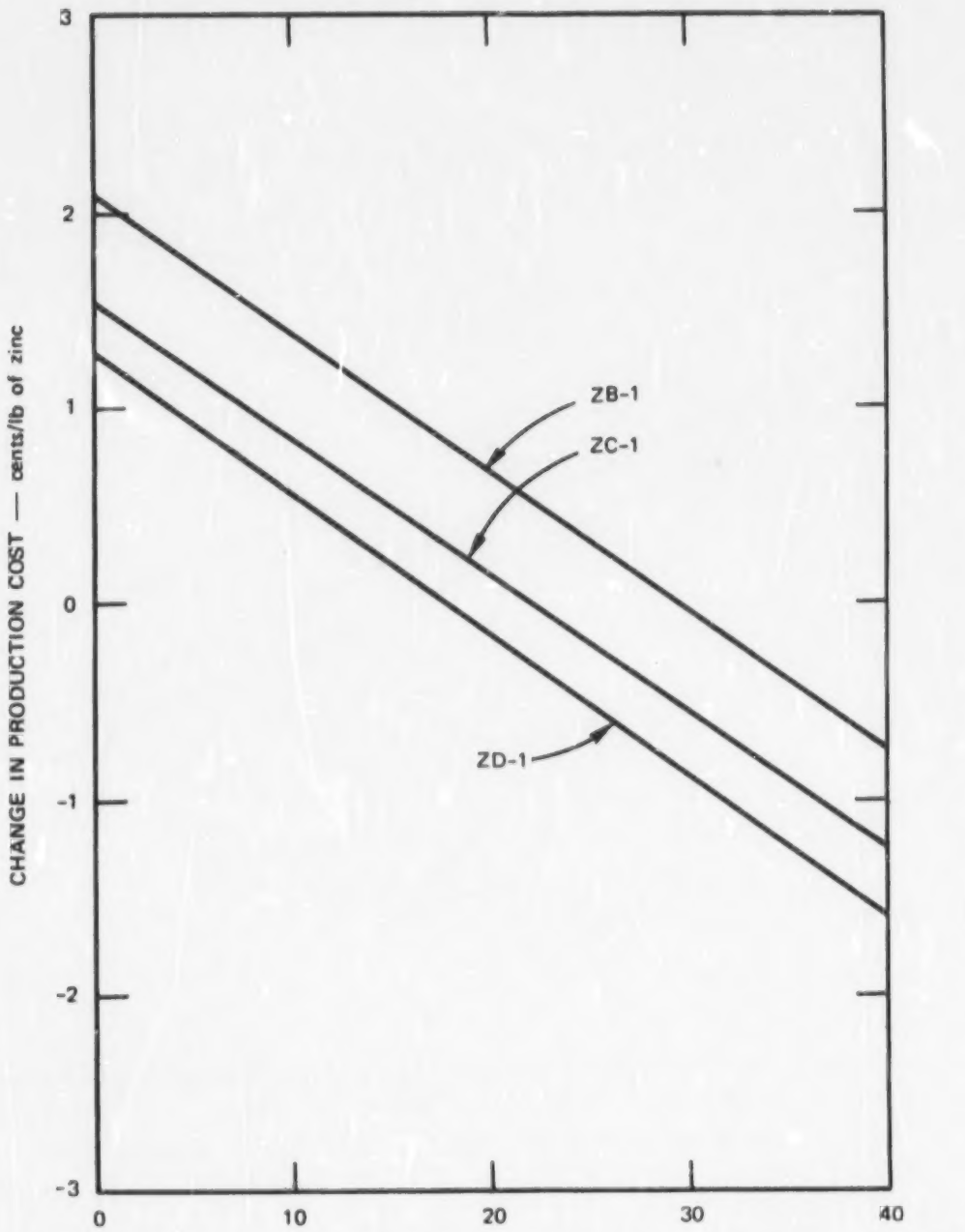

FIGURE 11 ZINC SMELTER MODELS B, C, AND D: CONTROL SCHEMES ZB-1, ZC-1, AND ZD-1—EFFECTS ON ZINC PRODUCTION COSTS

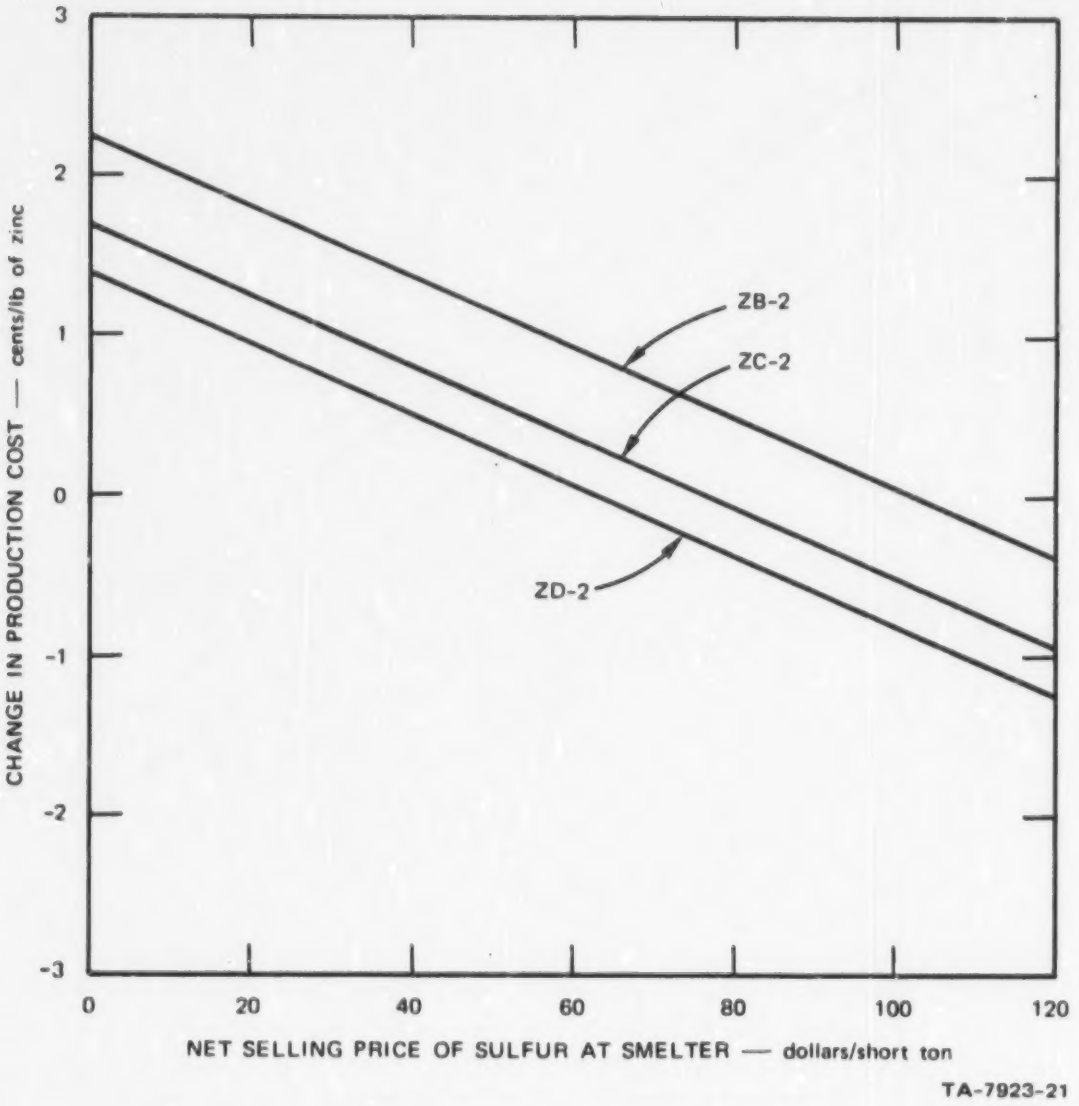

FIGURE 12 ZINC SMELTER MODELS B, C, AND D: CONTROL SCHEMES ZB-2, ZC-2, AND ZD-2—EFFECTS ON ZINC PRODUCTION COSTS

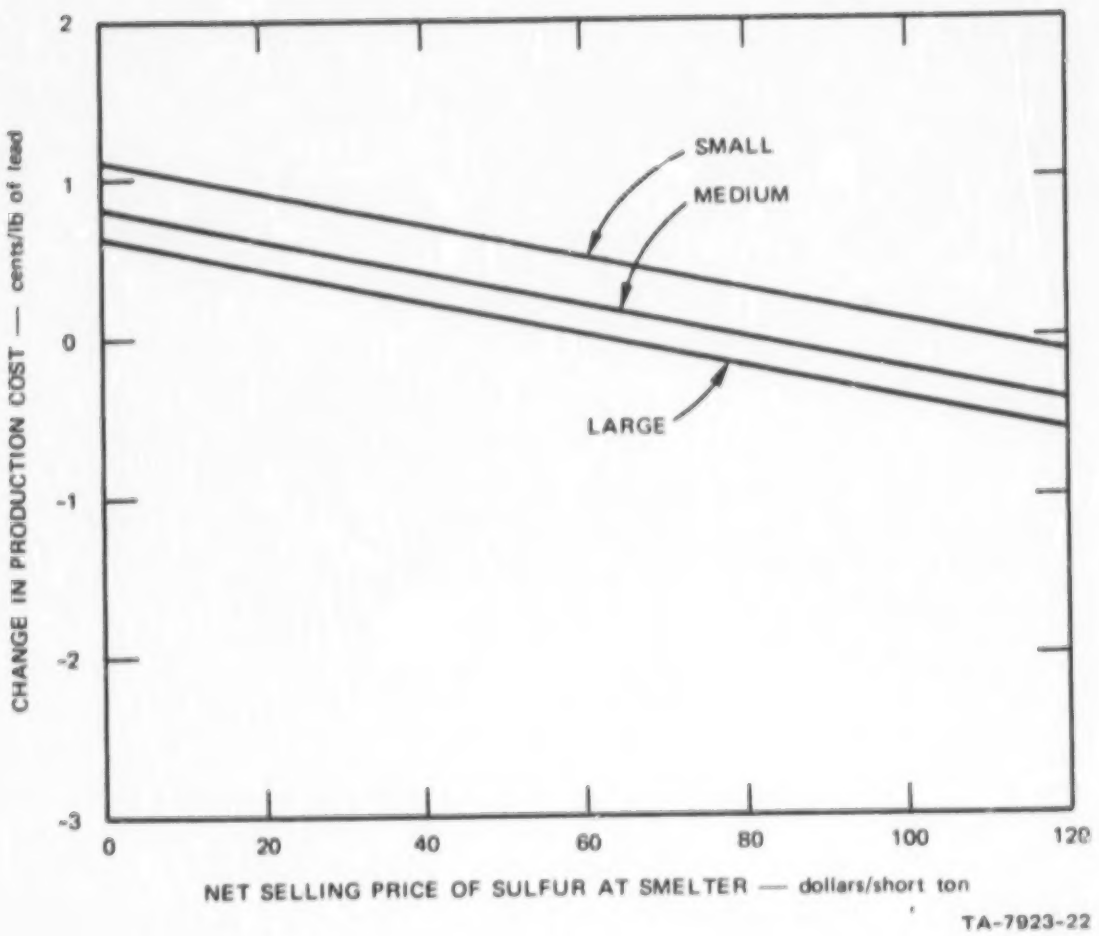

FIGURE 13 MODEL A LEAD SMELTERS: CONTROL SCHEME LA-1—EFFECT ON LEAD PRODUCTION COSTS

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VII GENERAL DISCUSSION

A. Evaluation of the Wellman-Lord SO_2 Recovery System

The Wellman-Lord process using sodium-base absorbent, described above, appears to be technically feasible. The regeneration of the absorbent, which is, in the writer's opinion, the only really novel aspect of the process, appears to be feasible and to offer a significant reduction of the steam consumption from the levels required by direct steam stripping of the rich absorbent. The capital investment in the Wellman-Lord chemical recovery system is probably much higher than that for a simple steam-stripping system. However, the associated fixed charges on the higher investment should be more than compensated by the reduction of twofold or threefold in the steam requirement.

A major unresolved question concerning the Wellman-Lord system is the amount of oxidation of the sulfite to sulfate. Some data should be acquired during 1970 from the commercial plant being constructed for use on sulfuric acid plant tail gas, but they are unlikely to be generally applicable even when they do become available. The general effectiveness of oxidation inhibitors has not been established. There appears to be no way short of actual experience under the proposed operating conditions by which the amount of sulfite oxidation can be predicted with much confidence. Experience with other recovery systems[2,5,9,15,19] suggests that the amount of the sulfur dioxide oxidized might be in the range of 0.2 to 2.0 percent or even higher, dependent upon the conditions of the particular case.

Two of the most critical factors in the oxidation of the sulfur dioxide are the oxygen concentration in the gas[2,19] and the presence of oxidation catalysts in the absorbent solution.[10,11,21] In smelter gases, high oxygen concentrations are frequently present as the result of air infiltration. The high efficiency cleaning of the smelter gas ahead of the absorber will be helpful because it will reduce the

introduction of potential oxidation catalysts into the absorbent. However, some of the catalysts are effective at very low concentrations.[10,11]

As is discussed above in Section V, oxidation of as little as 4 to 6 percent of the sulfur dioxide can result in loss of a substantial fraction of the total sulfur dioxide during purging of the sulfate. Therefore, there is potentially a need for an alternative method for separating the sodium sulfate from the sodium sulfite so that most of the latter can be returned as such to the absorption cycle.

The type of absorber proposed by Wellman-Lord is neither the only, nor necessarily the best, that might be employed. Stone & Webster Engineering Corp.[20] prepared estimates of the costs of various types of absorbers for use with the same sodium sulfite-bisulfite system on applications to power plant flue gases. Except for the factor of cost, however, Stone & Webster was unable to present any information that provides a clear choice between absorber designs.

For smelter applications, the cost of the Wellman-Lord system can be divided into three parts corresponding to the gas cleaning section, the absorption section, and the chemical recovery section. (The costs of the gas cleaning section would be common to any other recovery system.) The costs of the gas cleaning section and the absorption section are related to the volume of gas, but the cost of the chemical recovery section is related only to the quantity of sulfur dioxide handled. The costs of the auxiliary conversion plants (contact acid or sulfur dioxide reduction) are, of course, related to the quantity of sulfur dioxide; these costs, too, would be common to other sulfur-dioxide concentration processes.

The estimated costs for the various model control schemes show clearly the trends in cost distribution that are related to sulfur dioxide concentration in the gas. For sulfur dioxide concentrations below 1 0 percent, the capital cost of the gas cleaning section is the largest item, with that of the chemical recovery section being second and that of the absorption section much lower. In annual cost, however, the chemical recovery section is the largest item. With a sulfur dioxide concentration of about 5 percent

or higher, the chemical recovery section is by far the largest item in both capital and annual cost. As the sulfur dioxide concentration is increased above the range of 5 to 7 percent, the proportionate reduction in cost (for a given quantity of sulfur dioxide handled) will be relatively small, because the costs of the gas handling sections are already a minor component, and the cost of the chemical recovery section remains the same.

This trend in costs is inherent in the use of sodium sulfite (or sulfites of other strong bases) as an absorbent. The equilibrium partial pressure of sulfur dioxide is favorable to absorption of sulfur dioxide from lean gases. However, the capacity for sulfur dioxide is fixed by the stoichiometric quantity of reactant, and the quantity of absorbent used has to remain essentially proportional to the total quantity of sulfur dioxide regardless of concentration. The Wellman-Lord system -- as well as others using the same class of absorbents -- is primarily suited for use on gases containing up to about 3 percent, or perhaps 4 percent, of sulfur dioxide. For treatment of richer gases, a more favorable absorbent is one in which the sulfur dioxide is physically absorbed, or present only in a relatively weak chemical combination, and the capacity of the solvent is markedly a function of the partial pressure of the sulfur dioxide. Dimethylaniline is an example of a solvent for which the sulfur dioxide capacity continues to increase as the sulfur dioxide concentration in the gas rises to at least 8 or 9 percent.[9] With such a solvent, the quantity that must be circulated and the quantity of steam required for regeneration, per unit quantity of sulfur dioxide recovered, will continue to decrease as increasingly rich gas is treated, making the recovery system more economic.

B. Disposal of By-Products

The sizes and availabilities of markets for sulfur by-products producible at smelters have been presented in the McKee report,[16] in which estimated by-product prices were based on an assumed Gulf Coast sulfur price of $30/long ton, f.o.b. At the time of the previous study,[16]

it was estimated that the price of $30 was a likely average for the period up to 1975. However, sulfur supplies have since shifted from shortages to surpluses in a period of only about a year, and sulfur prices have become chaotic. The development of sulfur surpluses has followed a continuing period of low activity in the market for fertilizers, which provide the largest outlet for sulfur.

In the next five years the withdrawal of marginal producers and a possible revival of the fertilizer market may cause sulfur prices to rise from their present lows. However, the figure of $30/long ton now appears to be an optimistic one. It perhaps represents the upper end of the probable range of Gulf Coast prices to be encountered in the period to 1975. It cannot even be assumed that the Gulf Coast sulfur price will continue to maintain its previous status as the base line for world sulfur prices. Nevertheless, if the above limitations are recognized, the assumption of the $30/long ton price is probably as good as any that can be made at this time. The estimates of sulfur by-product prices in smelter areas that are presented in Reference 16 can therefore be used as a first approximation if it is understood that they probably represent the most favorable situation that can be anticipated.

Of those model control schemes that produce sulfuric acid, only Control Scheme CB-1 gives an estimated acid production cost low enough possibly to be competitive in those areas of the United States where there are a large number of uncontrolled smelter emissions: (1) Idaho and Montana, and (2) Arizona, New Mexico, and western Texas. Estimated acid production costs for all the other control schemes are clearly too high to be competitive.

Elemental sulfur is a far preferable product, because it can be shipped much greater distances for sale than can acid, and it can be easily stored. However, the estimated production costs of sulfur from all of the control schemes are far higher than the prices that were generally obtainable even during the sulfur shortage that has just passed.

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22. Terrana, J. D., and L. A. Miller, Process for Recovery of Sulfur Dioxide from Stack Gases, Proc. Amer. Power Conf. 30, 627-632 (1968) (Pub. 1969)

23. Watt, Stuart G., Wellman-Lord SO_2 Recovery Process, Wellman-Lord, Ind., Lakeland, Florida (1969)

24. Yodis, A. W., Applicability of Reduction to Sulfur Techniques to the Development of New Processes for Removing SO_2 from Flue Gases -- Phase I. Allied Chemical Corp., Interim Report to National Air Pollution Control Administration for Period June 1, 1968-July 31, 1969 (Draft Copy, Sept. 26, 1969), Contract No. PH 22-68-24.

Appendix A

SMELTER MODELS

The models of hypothetical copper, lead, and zinc smelters were created by Arthur G. McKee & Company, and are described in full in the McKee report.[A2] The essential features of the models directly pertinent to the present study are summarized here. Some additional model conditions, specific to the present study, were set by SRI, as indicated below.

In Table A-1 are presented the data on gas stream compositions, gas flow rates, and sulfur emissions for each of the model smelters. The data on metal production at the model smelters are given in Table A-2. The remainder of the model conditions are summarized below. Conditions specified by SRI are noted by an asterisk.

A. Plant Operating Time

330 days/year

24 hours/day

B. Gas Cleaning

1. Gases are assumed to have been cleaned in a hot electrostatic precipitator to a residual dust and fume content of 0.1 grain/std cu ft. The cost of hot, dry gas cleaning is not charged against sulfur oxides recovery.

*2. The gas leaving the hot precipitator is assumed to be cleaned in a system typical of those used for cleaning feed gases to a contact sulfuric acid plant[A1] (see Appendix C). The costs (capital and operating) of this secondary cleaning system are charged against sulfur oxides recovery. The cleaned gas is assumed to be at 110°F and essentially 1 atm pressure, and to be saturated with water vapor. It is further assumed that the residual concentrations of impurities are at least as low as those Donovan and Stuber[A1] have specified to be acceptable in contact sulfuric acid plants:

Table A-1

SMELTER MODELS
SUMMARY OF GAS COMPOSITIONS AND FLOW RATES AND OF SULFUR EMISSIONS

Plant	Model	Process Unit	Gas Temp. (°F)	Gas Composition (%)				Gas Flow Rate 1,000 SCFM (32°F, 1 atm)			Sulfur Equivalent (short tons/day)		
				SO_2	SO_3	O_2	H_2O	Small	Medium	Large	Small	Medium	Large
Copper	A	Reverberatory Furnace	550	1.89		6.48	11.7	69.9	139.8	209.7	84.6	169.2	253.8
		Converters	645	3.82				64.8 [1]	129.6 [2]	194.4 [2]	158.9	317.8	476.7
	B	Roaster	550	8.00		5.71	34.8	36.0	72.0	108	183.8	367.6	551.4
		Reverberatory Furnace	550	0.91				42.15	84.3	126.5	24.6	49.2	73.8
		Converters	637	3.78		16.1		58.04 [3]	116.1 [4]	174.1 [4]	141.6	283.2	424.8
Zinc	A	Roaster	600	7.1	0.1	10.9	0.05	19.31	38.62	57.93	89.6	179.2	268.8
		Sinter plant	500	0.048	0.0016	18.0	3.2	29.8	59.6	89.4	0.9	1.8	2.7
	B	Roaster	600	0.9		18.0		133.2	--	--	77.1	--	--
		Sinter plant	300	0.5		18.0		47.9	--	--	15.4	--	--
	C	Roaster	400	0.8		16.0		--	353.4	--	--	181.8	--
	D	Sinter-Roaster	300	2.0		16.0		--	144.0	--	--	185.2	--
Lead	A	Sinter plant	400	5.0	0.29	12.0		10.67	21.34	42.68	34.3	68.6	137.2

[1] Size recovery plant for 137% of average flow.
[2] Size recovery plant for 130% of average flow.
[3] Size recovery plant for 130% of average flow if gases combined with roaster gases.
[4] Size recovery plant for 123% of average flow if gases combined with roaster gases.

Table A-2

METAL PRODUCTION BY MODEL SMELTERS

Metal	Model	Size	Metal Production tons/day	Metal Production tons/year
Copper	A	Small	230	75,900
		Medium	460	152,000
		Large	690	228,000
	B	Small	230	75,900
		Medium	460	152,000
		Large	690	228,000
Zinc	A	Small	171.4	56,600
		Medium	342.8	113,000
		Large	514.2	170,000
	B	Small	171.4	56,600
	C	Medium	342.8	113,000
	D	Medium	342.8	113,000
Lead	A	Small	142.9	47,200
		Medium	285.8	94,300
		Large	571.6	189,000

Contaminant	Concentration, grains/std cu ft, dry basis
Chlorides (as Cl)	0.0005
Fluorides (as F)	0.0001
Arsenic (as As_2O_3)	0.0005
Lead (as Pb)	0.0005
Mercury (as Hg)	0.0001
Selenium (as Se)	0.022
H_2SO_4 mist (100%)	0.022
Total Solids (dust)	0.0005

*C. Sulfur Dioxide Collection Efficiency

The performance of the primary sulfur dioxide collection system is to be as follows, defined in terms of collection efficiency or of sulfur dioxide concentration in the exit gas:

SO_2 Concentration in Feed Gas (%)	Efficiency (%) or Concentration in Exit Gas (ppm)
Under 1.0	500 ppm
1.0 to 2.0	95%
Over 2.0	98%

In cases where sulfur dioxide is concentrated for conversion to another product in a secondary process, the efficiency of the conversion step is to be taken as a reasonable one for the process assumed.

*D. Cooling Water

The capital and operating costs of the sulfur dioxide control system are to include the costs of a water cooling tower. Cooling water costs will be for makeup only.

References

A1. Donovan, J. R., and P. J. Stuber, Sulfuric Acid Production from Ore Roaster Gases, J. Metals 19 (11), 45-50 (Nov. 1967)

A2. McKee & Company, Arthur, G., Systems Study for Control of Emissions -- Primary Nonferrous Smelting Industry, Final Report to National Air Pollution Control Administration, June 1969, Contract No. PH 86-68-85.

Appendix B

FACTORS AND CONDITIONS ASSUMED IN ESTIMATING CONTROL SYSTEM COSTS

The following conditions and cost factors were taken from the report by Arthur G. McKee & Company,[B1] except that the items marked with an asterisk are additional conditions specified by SRI.

A. Capital Costs

1. Capital costs do not include:

   (a) Working capital
   (b) Contingencies
   (c) Cost of land
   (d) Inventory
   (e) Interest on investment
   (f) Offsite utilities
   (g) Steam generators
   (h) Plant access
   (i) Cost of dry, hot gas cleaning

2. The cost of the wet cleaning system for the gas is charged to sulfur oxides control, but is estimated as a separate item (see Appendix C).

B. Depreciation

15-year, straight-line.

C. Direct Operating Costs (except utilities)

1.	Labor	$3.75/man-hour
2.	Supervision	4.75/man-hour
3.	Payroll benefits	25% of labor + supervision
4.	Maintenance materials	3% of capital cost of plant
5.	Factory supplies	0.5% of capital cost of plant

D. Indirect Costs

1. Controllable indirect costs (maintenance labor, laboratory, overhead, and supervision)

   50% of labor + supervision + maintenance materials

2. Noncontrollable indirect costs (local taxes and insurance)

   3% of capital cost of plant

E. Utilities

1.	Electrical power	1¢/kwh
2.	Steam	80¢/1000 lb *(400 psig, 750°F)
3.	Natural gas	40¢/million Btu
4.	Water	
	(a) Makeup cooling water	2¢/1000 gal
	(b) Process water	20¢/1000 gal
	*(c) Boiler feed water	40¢/1000 gal
	*(d) Retreatment of process steam condensates for boiler use	38¢/1000 gal
	*(e) Cooling water temperature	80°F

F. Waste Acid Disposal

(Weak, waste acid from preliminary gas cooling and cleaning.)
Cost of weak acid disposal 50¢/1000 gal

G. Plant Operating Time

330 days/year
24 hours/day

Reference

B1. McKee & Company, Arthur G., Systems Study for Control of Emissions -- Primary Nonferrous Smelting Industry, Final Report to National Air Pollution Control Administration, June 1969, Contract No. PH 86-68-85.

Appendix C

GAS CLEANING SYSTEM

The flowsheet for the gas cleaning system (Fig. C-1) was adapted from that portrayed for contact sulfuric acid plants in the Arthur G. McKee & Company final report.[C2] The inlet gas conditions were also assumed to be the same as those specified in the McKee report, except that a single, average value was taken for the SO_3 concentration (0.2% by volume). Under the conditions assumed, the water vapor condensed from the incoming gas is sufficient to balance the outflow of weak acid waste, so that no makeup scrubbing water is required. The cooling water used in the heat exchanger to cool the recirculated scrubbing water is assumed to be itself cooled in a water cooling tower. The water cooling tower is not shown in the flowsheet but is assumed to be part of the system. It is treated as a separate unit for convenience in cost accounting, even though the same water cooling tower would probably in most cases handle the cooling water for both the gas cleaning and the sulfur dioxide recovery systems.

The capital cost for the gas cleaning system (not including water cooling tower) was taken from the paper by J. M. Connor.[C1] Data points were taken from cost curve No. 2 of Connor's Fig. 5, corresponding to indirect cooling with a $30^{o}F$ approach between the cold gas and cooling water temperatures. The system capacity was converted from the original basis (tons per day of 100% H_2SO_4 with gas containing 8% SO_2) to the basis of volumetric gas flow at system exit conditions ($110^{o}F$, 1 atm, sat'd).

Connor's estimates were presumably made for the type of gas cleaning system illustrated in his paper, a scrubber-cooler followed by an indirect tubular type of gas cooler, with an electrostatic mist precipitator for final cleanup. Although the scrubbing and cooling equipment may therefore be different from that assumed in the present case, Connor's cost estimates (which were admittedly approximate) were taken to be sufficiently precise

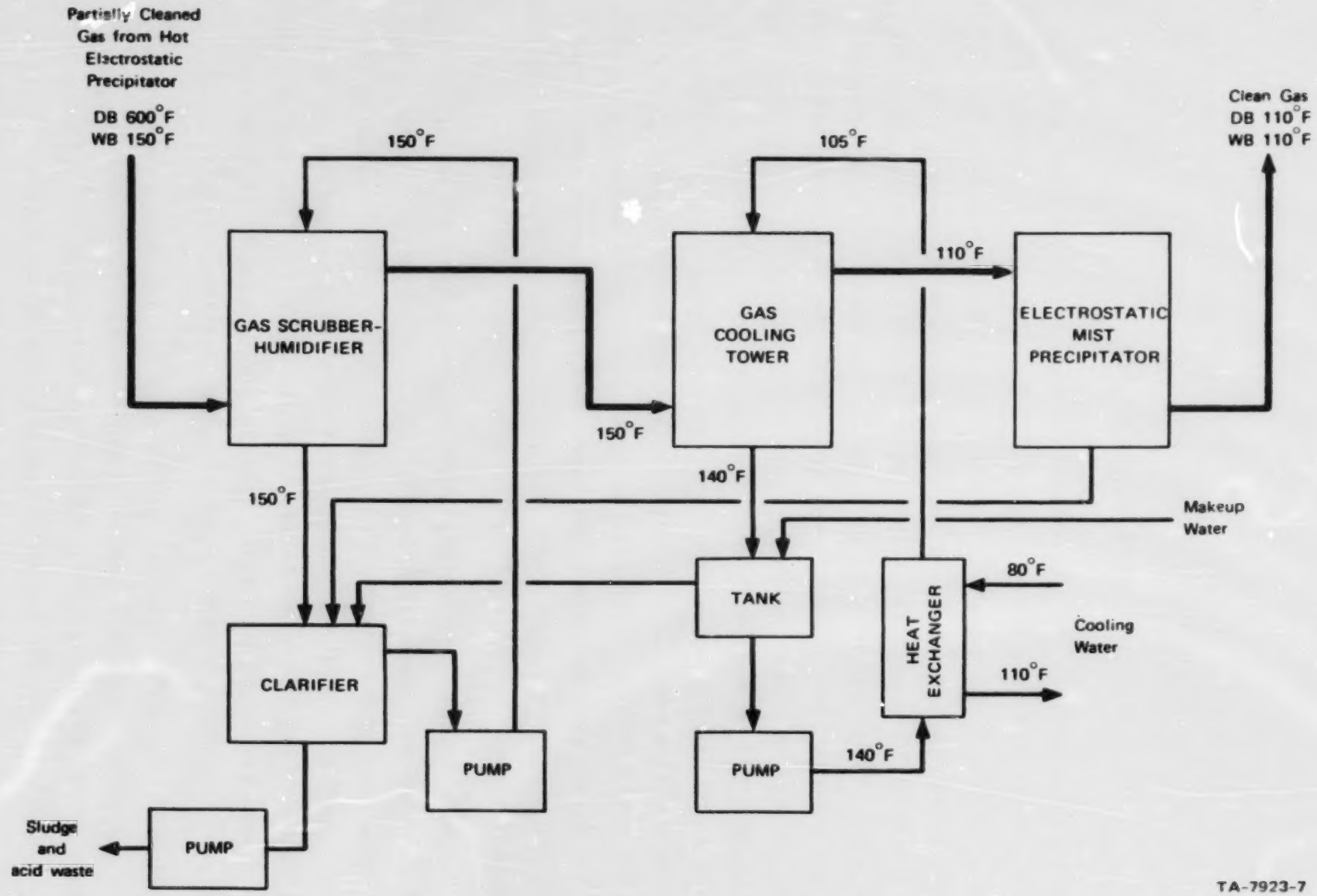

FIGURE C-1 GAS CLEANING SYSTEM FOR SMELTER GASES

for use in this analysis.

Connor's curves did not extend beyond a capacity equivalent to about 100,000 CFM of gas, but the plotted curve (Fig. C-2) was extrapolated arbitrarily to 1,000,000 CFM. The actual upper limit of validity probably corresponds to no more than 300 to 400,000 CFM. Above this size range, the use of multiple units of equipment may make the cost more nearly proportional to the first power of gas flow rate. However, none of the individual gas flows in the smelter models exceeds about 400,000 CFM, above which the curve in Fig. C-2 is shown as a broken line.

To obtain the curve of capital costs (Fig. C-2) for use in the present study, the capital cost of a water cooling tower was added to the cost of the remainder of the system as adapted from Connor's data.

The gas and water flows through the system were determined by making heat and material balances based on the conditions shown in Fig. C-1. Power requirements were estimated from literature data in a few instances, and calculated from guesstimates of gas pressure drop and pump heads in the rest. Utilities, labor requirements, and cost factors employed are summarized in Tables C-1 and C-2. The standard cost factors used were the same as those employed elsewhere in the study (Appendix B).

The total annual costs were calculated for five gas flow capacities ranging from 10,000 to 1,000,000 CFM (see Table C-3), using the data of Fig. C-2 and Tables C-1 and C-2, and were used to prepare the curve of annual cost, Fig. C-3. The curve probably is not valid for capacities greater than 300 to 400,000 CFM, and is shown as a broken line above the latter level.

References

C1. Connor, J. M., Economics of Sulfuric Acid Manufacture, Chem. Eng. Progr. 64 (11), 59-65 (Nov. 1968)

C2. McKee & Company, Arthur G., Systems Study for Control of Emissions -- Primary Nonferrous Smelting Industry, Final Report to National Air Pollution Control Administration, June 1969, Contract No. PH 86-68-85.

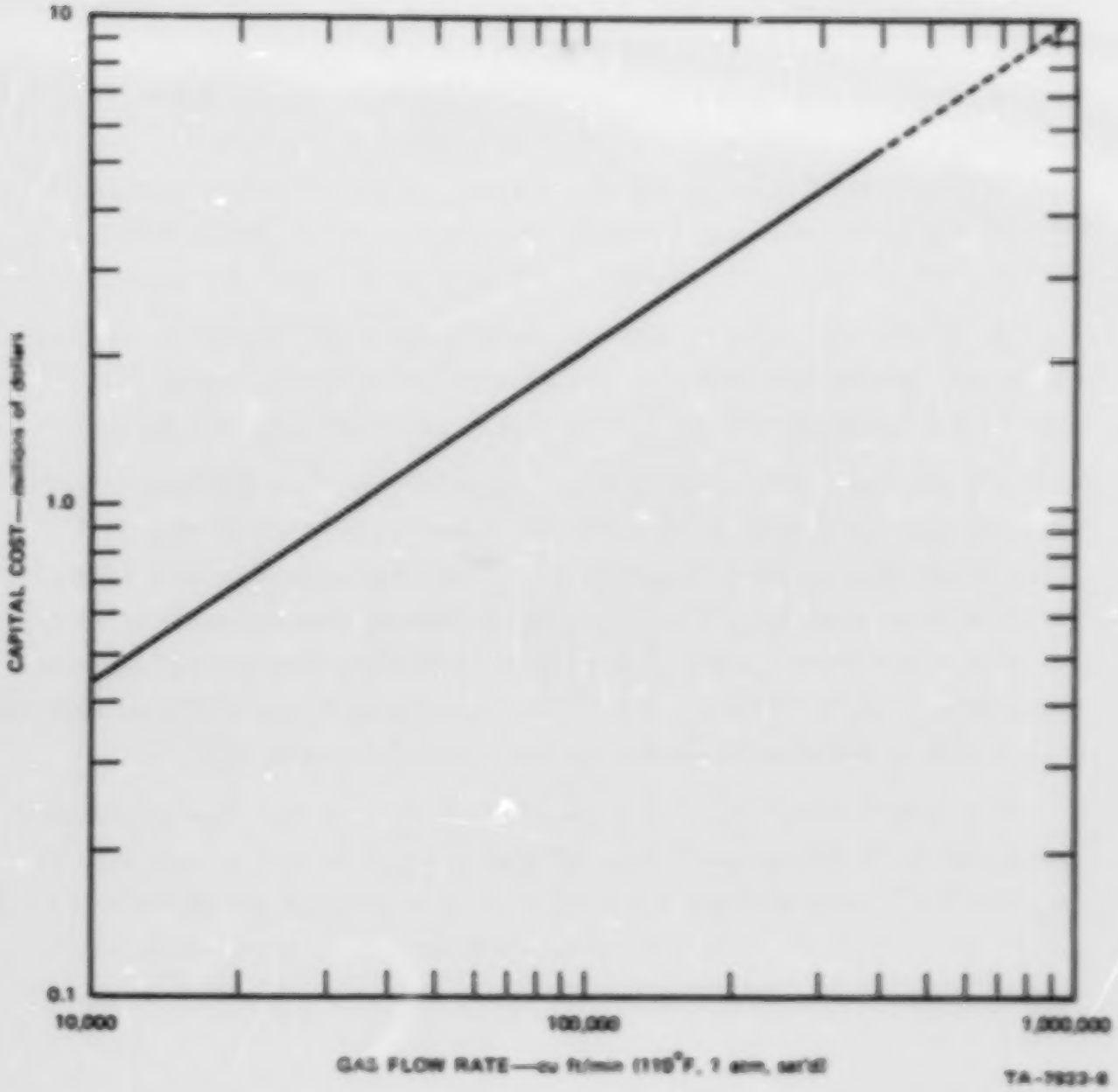

FIGURE C-2 CAPITAL COST OF GAS CLEANING SYSTEM FOR SMELTER GASES

Table C-1

GAS CLEANING SYSTEM FOR SMELTER GAS
LIQUID FLOWS AND POWER REQUIREMENTS

Item	Basis Gas Flow = 1000 CFM at 110°F, 1 atm, sat'd	
	Liquid Flow	Power
Gas Flow through System Δp = 14 in. w.c.		3.50 hp
Scrubber-Humidifier Circulated water	7.86 gpm	0.43 hp
Gas Cooling Tower Circulated water	35.7 gpm	1.50 hp
Cooling Water Circulation	41.7 gpm	1.40 hp
Waste Acid (21% H_2SO_4)	0.212 gpm	0.06 hp
Water Cooling Tower Circulated water Makeup water	41.7 gpm 4.17 gpm	1.91 hp
Subtotal		8.80 hp 6.57 kw
Electrostatic Precipitator		1.50 kw
Total		8.07 kw

Table C-2

GAS CLEANING SYSTEM FOR SMELTER GASES
UTILITIES AND OPERATING LABOR REQUIREMENTS

Item	Unit Cost	Utilities Basis:[1] Gas Flow = 1000 CFM @ 110°F, 1 atm, sat'd	
		Quantity	Annual Cost ($)
Electrical Power	$0.01/kwh		
Gas flow only		2.62 kw	207.50
Total		8.07 kw	639.14
Makeup Cooling Water	$0.02/1000 gal	4.17 gpm	39.63
Waste Acid Treatment	$0.50/1000 gal	0.212 gpm	50.37
Operating Labor	$3.75/hr	12 hrs/day[1]	14,850[1]
Supervision	$4.75/hr	2 hrs/day[1]	3,135[1]
Payroll Benefits	25% of Labor + Supervision		4,496[1]

Plant operating time = 330 days = 7920 hrs/yr.

[1]Labor costs are total for plant regardless of size.

Table C-3

GAS CLEANING SYSTEM FOR SMELTER GASES
CAPITAL AND TOTAL ANNUAL COSTS

Item	Cost Basis	Gas Flow Rate-CFM at 110°F, 1 atm, saturated				
		10,000	30,000	100,000	300,000	1,000,000
Capital Investment	$	436,000	930,000	2,104,000	4,480,000	10,156,000
	$/CFM	43.60	31.00	21.04	14.93	10.16
Annual Cost	$/yr					
A. Depreciation		29,100	62,000	140,200	298,670	677,100
B. Direct Operating Cost						
1. Labor		14,850	14,850	14,850	14,850	14,850
2. Supervision		3,135	3,135	3,135	3,135	3,135
3. Payroll benefits		4,496	4,496	4,496	4,496	4,496
4. Maintenance Materials		13,080	27,900	63,120	134,400	304,680
5. Factory supplies		2,180	4,650	10,520	22,400	50,780
6. Electricity		6,390	19,170	63,910	191,740	639,140
7. Makeup water		396	1,190	3,960	11,890	39,630
8. Waste acid treatment		504	1,510	5,040	15,110	50,370
C. Indirect Costs						
1. Controllable		15,530	22,940	40,550	76,190	161,330
2. Noncontrollable		13,080	27,900	63,120	134,400	304,680
Total Annual Cost	$/yr	102,741	189,741	412,901	907,281	2,250,191
	$/yr per CFM	10.27	6.33	4.13	3.02	2.25

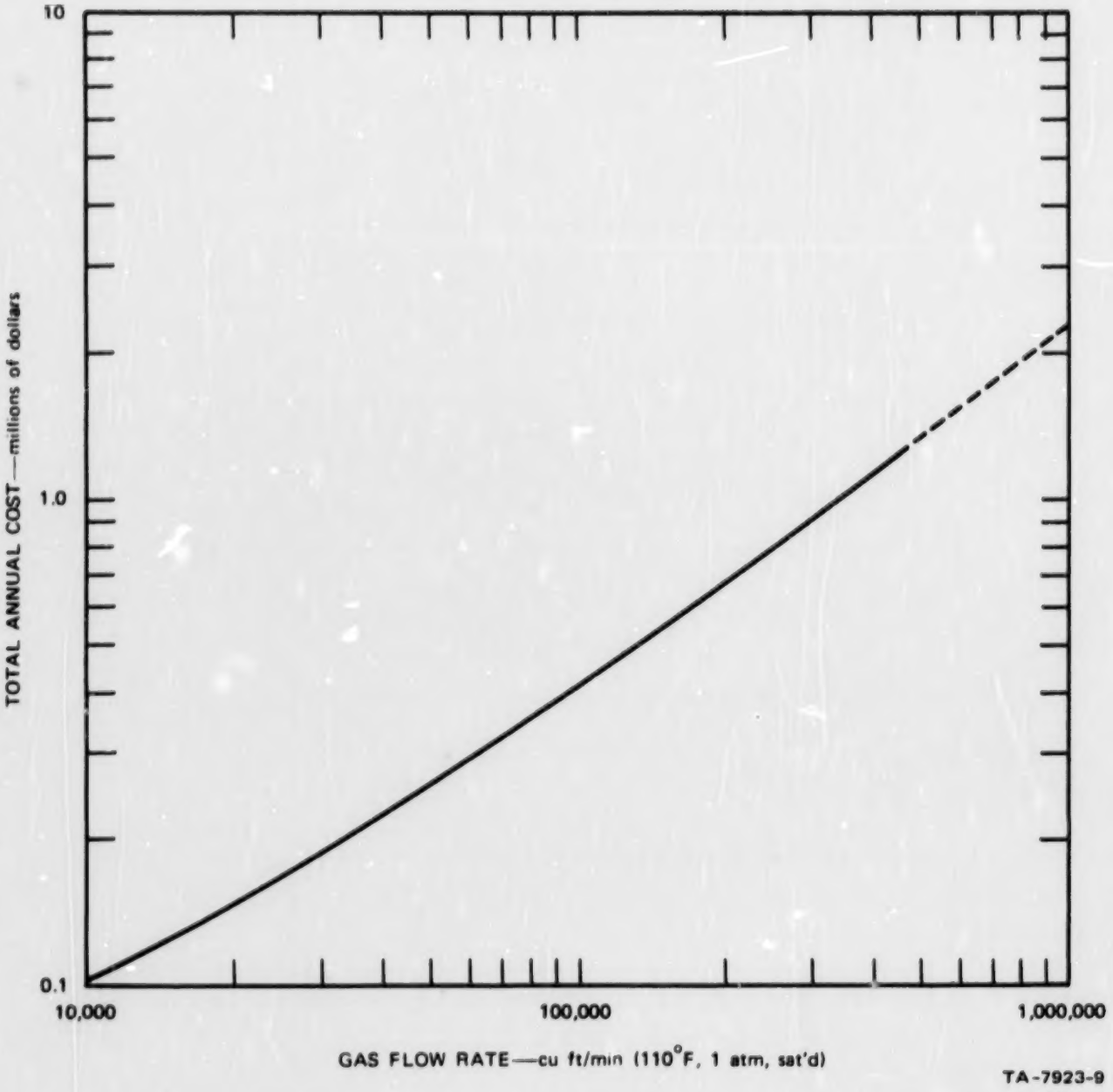

FIGURE C-3 TOTAL ANNUAL COST OF GAS CLEANING SYSTEM FOR SMELTER GASES

Appendix D

AUXILIARY CONTACT SULFURIC ACID PLANTS

Estimates of the capital costs of contact sulfuric acid plants for conversion of concentrated sulfur dioxide to sulfuric acid were derived from Fig. 4 of a paper by Connor.[D1] This figure* gives the cost of the acid-making section, only, of a conventional contact plant. This cost should correspond closely to that of the auxiliary plant, which has no gas-production section. Sufficient air was assumed to be added to the sulfur dioxide to give a mole ratio of oxygen to sulfur dioxide of 1.3.[D2] The resulting concentration of sulfur dioxide in the mixed feed gas was 13.9 percent. Conversion of the sulfur dioxide was assumed to be 98 percent.

Estimates of labor and utility requirements (Table D-1) were based generally on selected values from the references cited (D1 and D2). The costs for water were modified from those presented in Appendix B. The process water was assumed to be steam condensate or its equivalent. The cost for cooling water was taken to include capitalization of cooling towers, since a cooling tower was not included in the acid plant cost estimate.

The capital costs for the acid plants are presented in Fig. D-1. The estimated annual costs are calculated in Table D-2 and presented graphically in Fig. D-2.

*Note:

As published,[D1] Fig. 4 of Connor's paper is presented erroneously. The data given here were taken from the correct form of Fig. 4, which has not been published.

REFERENCES

D1. Connor, J. M., Economics of Sulfuric Acid Manufacture, Chem. Eng. Progra. 64 (11), 59-65 (Nov. 1968)

D2. Duecker, W. W., and J. K. West (Eds.), "The Manufacture of Sulfuric Acid," Reinhold Publishing Co., New York (1959)

Table D-1

LABOR AND UTILITIES REQUIREMENTS OF AUXILIARY SULFURIC ACID PLANTS

Item	Quantity per ton 100% H_2SO_4 produced	Unit Cost
Electricity	25 kwh	1.0¢/kwh
Cooling Water	7000 gal	4¢/1000 gal
Process Water	20 gal	40¢/1000 gal
Labor	24 man-hr/day[1]	$3.75/hr
Supervision	6 man-hr/day[1]	$4.75/hr

[1] Labor and supervision are daily totals independent of plant size.

Table D-2

CAPITAL AND ANNUAL COSTS OF AUXILIARY SULFURIC ACID PLANTS

Item	Cost Basis	Sulfuric Acid Produced-tons/day		
		150	500	1,200
Capital Investment	$	450,000	975,000	1,700,000
Annual Cost	$/yr			
A. Depreciation		30,000	65,000	113,330
B. Direct Operating Cost				
1. Labor		29,700	29,700	29,700
2. Supervision		9,410	9,410	9,410
3. Payroll benefits		9,780	9,780	9,780
4. Maintenance materials		13,500	29,250	51,000
5. Factory supplies		2,250	4,880	8,500
6. Electricity		12,380	41,250	99,000
7. Cooling water		13,860	46,200	110,900
8. Process water		400	1,320	3,170
C. Indirect Costs				
1. Controllable		26,300	34,180	45,060
2. Noncontrollable		13,500	29,250	51,000
Total Annual Cost	$/yr	161,080	300,220	530,850

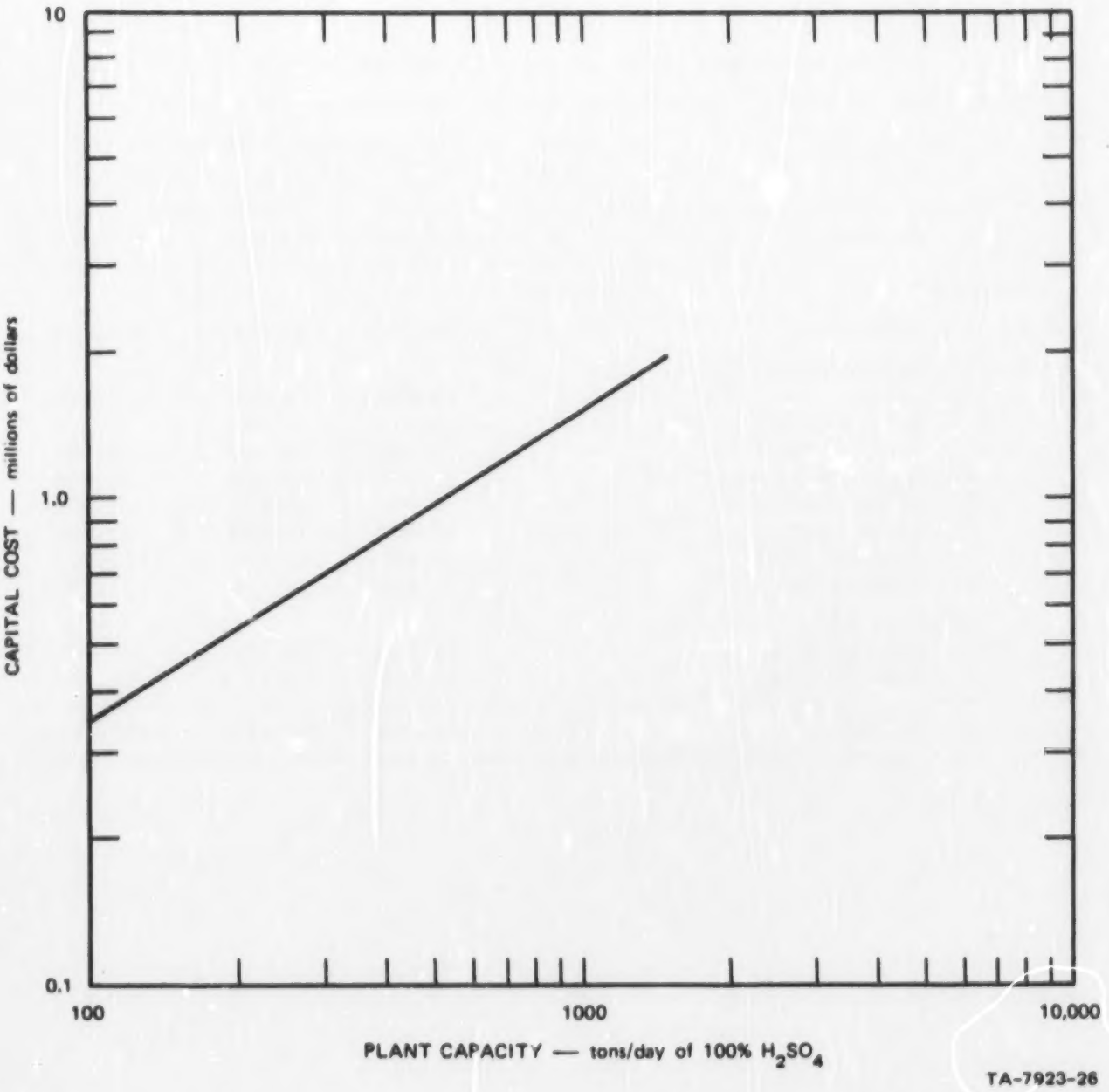

FIGURE D-1 CAPITAL COSTS OF AUXILIARY SULFURIC ACID PLANTS

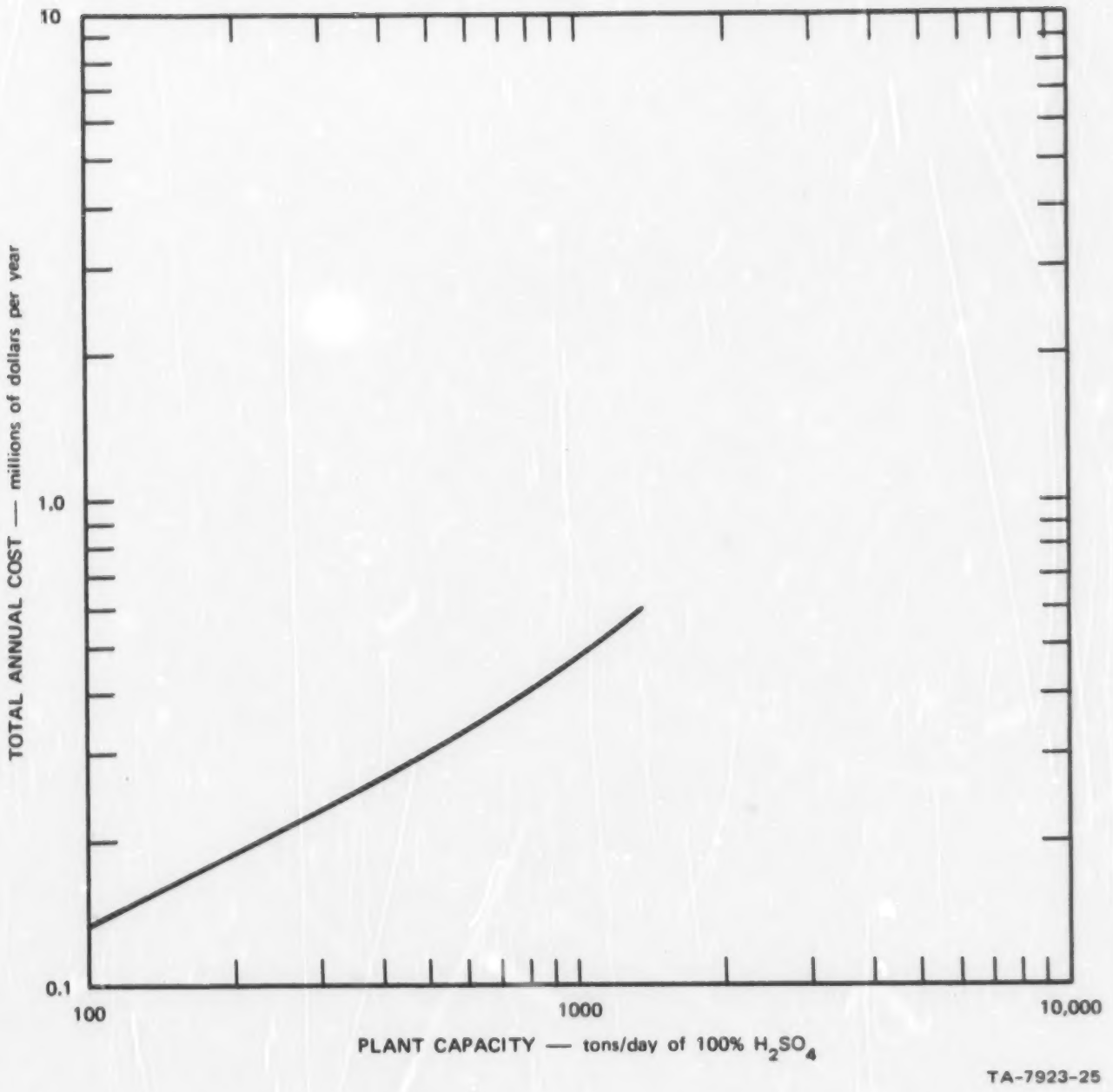

FIGURE D-2 ANNUAL COSTS OF AUXILIARY SULFURIC ACID PLANTS

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Appendix E

SULFUR DIOXIDE REDUCTION PLANTS

Estimates of the capital and operating costs of sulfur dioxide reduction plants were derived from data developed by Allied Chemical Corporation[E1] for a conceptual design based on the Asarco process. Allied presented costs for plants of two sizes, but the estimates for one were factored from detailed estimates made for the larger of the two. In Fig. E-1 the curve of capital cost was extrapolated (dashed line) to both smaller and larger sizes of plants; the validity of the extrapolation to the smaller sizes is probably the more questionable.

The estimates of labor, utilities, and raw material requirements (Table E-1) were taken from Reference E1. The cost factors were taken from Appendix B, with two exceptions. The cost of cooling water was taken to be a capitalized cost for cooling tower water, since the capital cost of the reduction plant did not include that for a cooling tower. The credit for low pressure steam was taken to be the same as used by Allied.

The yield of sulfur in the reduction plant was taken as 92.63 percent, as estimated by Allied.

The estimated annual costs for reduction plants are calculated in Table E-2 and are presented graphically in Fig. E-2.

REFERENCE

E1. Yodis, A. W., Applicability of Reduction to Sulfur Techniques to the Development of New Processes for Removing SO_2 from Flue Gases -- Phase I. Allied Chemical Corporation, Interim Report to National Air Pollution Control Administration for Period June 1, 1968 - July 31, 1969 (Draft Copy, September 26, 1969), Contract No. PH 22-68-24

Table E-1

LABOR, UTILITIES, AND RAW MATERIAL REQUIREMENTS OF SULFUR DIOXIDE REDUCTION PLANTS

Item	Quantity per 1000 lb SO_2 input	Unit Cost
Methane	2940 CF	40¢/1000 CF
Electricity	8.90 kwh	1.0¢/kwh
Cooling Water	420 gal	4¢/1000 gal
Boiler Feed Water Makeup	5 gal	40¢/1000 gal
Fuel Gas	170 CF	40¢/1000 CF
Catalyst	0.18 lb	13¢/lb
(Low Pressure Steam)[1]	(370 lb)[1]	(25¢/1000 lb)[1]
Labor	54 man-hr/day[2]	$3.75/hr
Supervision	8 man-hr/day[2]	$4.75/hr

[1] Credit.
[2] Labor and supervision are daily totals independent of plant size.

Table E-2

CAPITAL AND ANNUAL COSTS OF SULFUR DIOXIDE REDUCTION PLANTS

Item	Cost Basis	Sulfur Dioxide Input-lb/hr			
		3,000	10,000	30,000	200,000
Capital Investment	$	342,000	695,000	1,320,000	4,000,000
Annual Cost	$/yr				
A. Depreciation		22,800	46,330	88,000	266,700
B. Direct Operating Cost					
1. Labor		66,830	66,830	66,830	66,830
2. Supervision		12,540	12,540	12,540	12,540
3. Payroll benefits		19,840	19,840	19,840	19,840
4. Maintenance materials		10,260	20,850	39,600	120,000
5. Factory supplies		1,710	3,480	6,600	20,000
6. Electricity		2,110	7,050	21,140	140,960
7. Methane		27,940	93,140	279,420	1,862,800
8. Fuel gas		1,620	5,390	16,170	107,800
9. Cooling water		400	1,330	3,990	26,600
10. Boiler feed water		50	160	470	3,160
11. Catalyst		560	1,850	5,560	37,060
C. Indirect Costs					
1. Controllable		44,820	50,110	59,490	99,690
2. Noncontrollable		10,260	20,850	39,600	120,000
Gross Total Annual Cost	$/yr	221,740	349,750	659,250	2,903,980
Steam Credit	$/yr	2,200	7,330	21,980	146,520
Net Total Annual Cost	$/yr	219,540	342,420	637,270	2,757,460

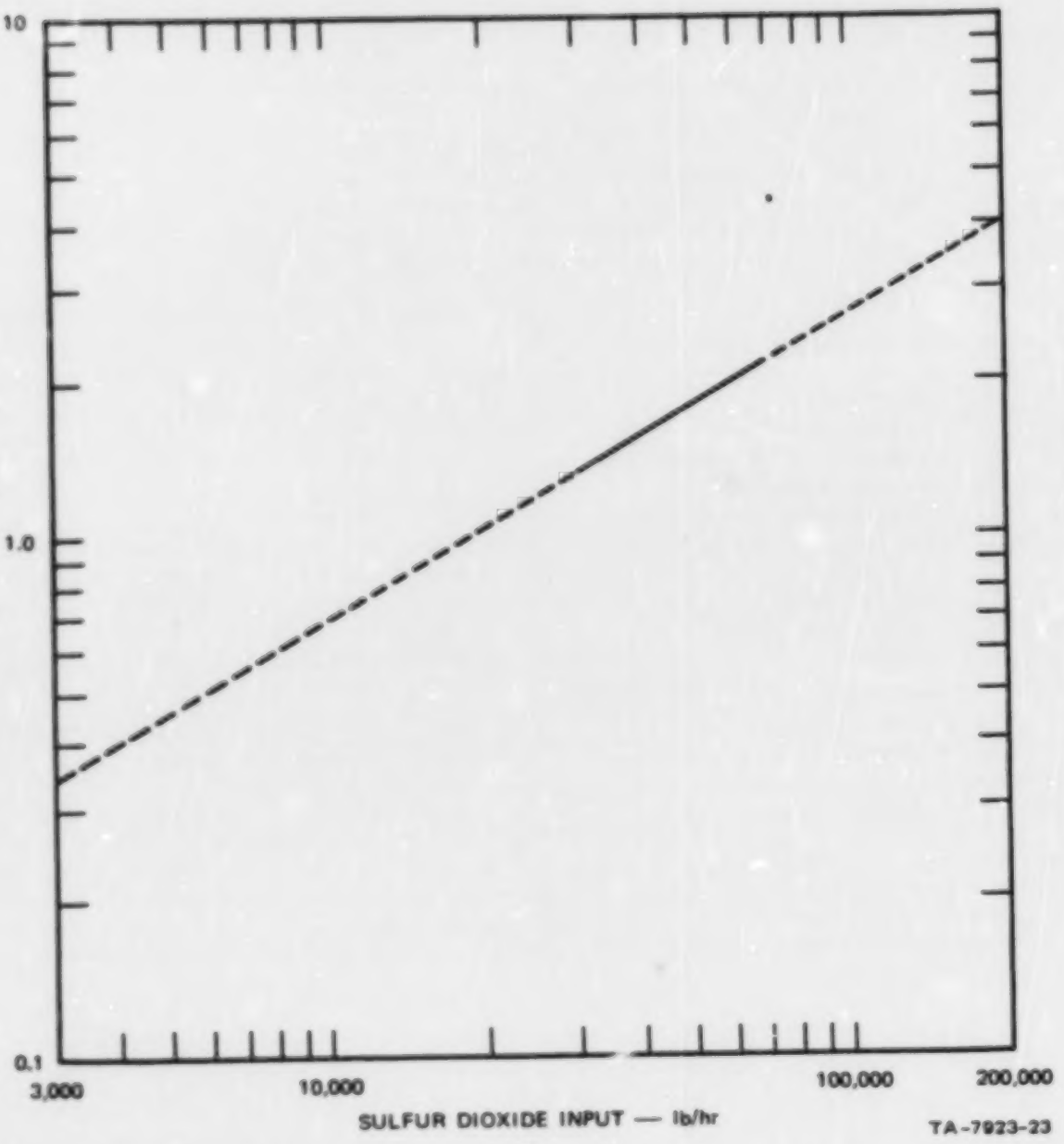

FIGURE E-1 CAPITAL COSTS OF SULFUR DIOXIDE REDUCTION PLANTS

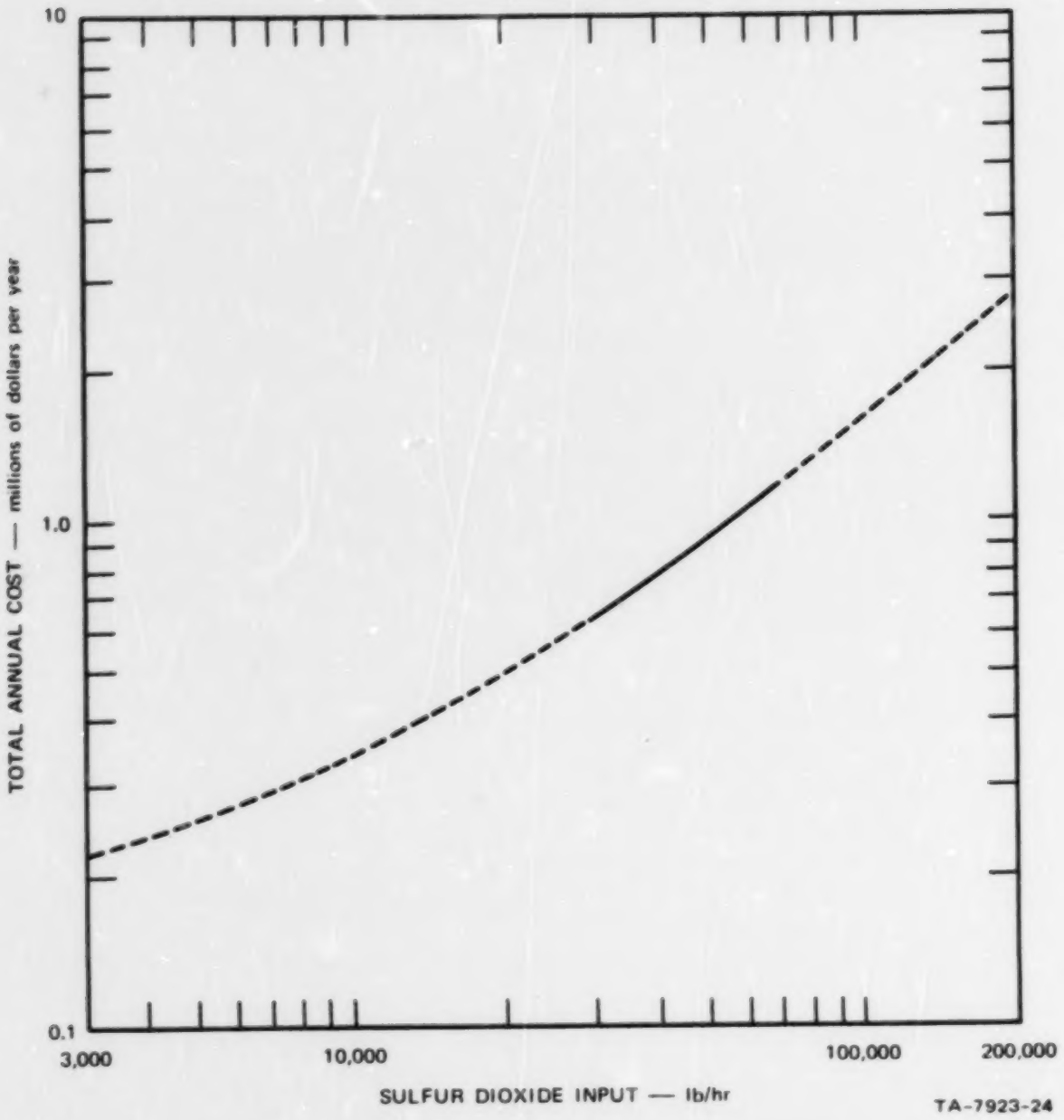

FIGURE E-2 ANNUAL COSTS OF SULFUR DIOXIDE REDUCTION PLANTS